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Flue Gas Desulfurization Systems Design and Operating Considerations
Volume II. Technical Report

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FLUE GAS DESULFURIZATION SYSTEMS: DESIGN AND OPERATING CONSIDERATIONS

Volume II. Technical Report

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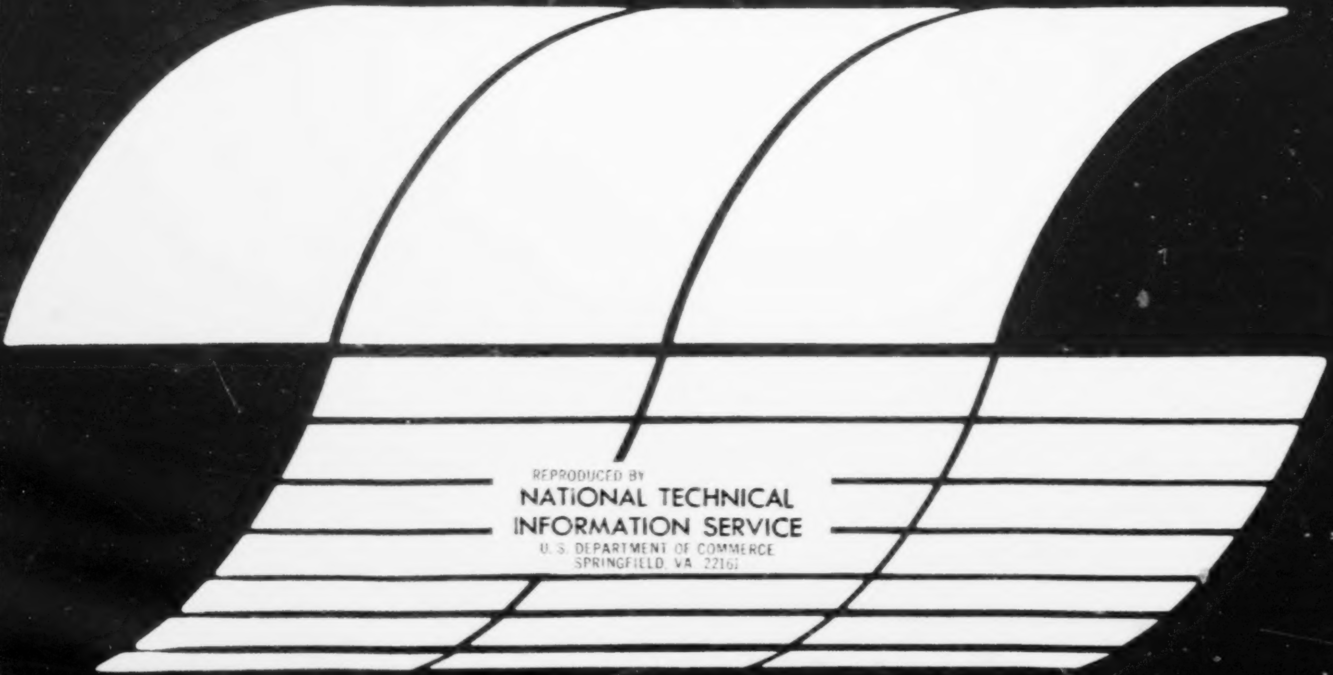

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16. ABSTRACT
The report describes flue gas desulfurization (FGD) systems and the design and operating parameters that are monitored to ensure proper operation. It explains how these parameters are varied to accommodate changing boiler loads and fuel characteristics, and describes the control of parameters to prevent such problems as scale buildup. It describes effects of designing and operating FGD systems for 90% or greater SO2 removal efficiencies, based on current testing program data. It describes effects of coal characteristics on FGD performance, along with operating and design techniques used to compensate for coal property variations. It describes the purpose, need, and methods for exhaust gas reheat, downstream of FGD systems. It discusses alternatives to exhaust gas reheat.

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Coal	Sulfur Dioxide	Particulate	21D	
Reheating	Dust	Venturi/Spray Towers	13A	11G
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FOREWORD

EPA has undertaken a review of the New Source Performance Standards for SO_2 emissions from coal-fired steam generators. As part of this review, EPA is assembling information related to the SO_2 removal capability of FGD systems and technical and economic data for the design of FGD systems for higher SO_2 removal efficiencies.

In the conduct of this review the EPA has contracted for the services of a number of supporting organizations, each of which will make specific contributions to the total subject. In this respect, a substantial number of separate, but related, issues will be investigated, including such broad issues as economic, energy, social, technological, and environmental issues.

This report addresses itself to the following specific tasks:

(1) Describe FGD system parameters which are commonly monitored to ensure proper operation and explain how those parameters are varied to accommodate combustion system variations such as changing boiler loads, changes in SO_2 inlet concentrations, and changes in other fuel characteristics. Describe the parameters which are controlled to limit scale build-up in the scrubber system and mist eliminator and the ranges within which they must be controlled and parameters which are controlled to limit other problems.

(2) Where an installed full-scale FGD system has demonstrated good operation, but at lower SO_2 removal efficiencies (such as 60 to 85 percent), define the changes that would be applied to a new facility to obtain 90 percent or greater SO_2 removal and present data showing the effects of these changes. For example,

SO_2 removal efficiency may be increased by increasing the liquid-to-gas ratio, changing pH, addition of MgO to lime/limestone scrubbers, changes in scrubber gas velocity, increasing bed heights in absorption towers, and others. Any rationale, engineering calculations, etc., necessary to extrapolate from an existing FGD system operating well at lower percent SO_2 removal to a 90 percent or greater SO_2 removal should be presented.

(3) Define the effects of changes in coal characteristics on FGD systems and describe operating parameters or design techniques which can be used to compensate for local and nationwide variations in coals. For example, coals with high chloride contents affect SO_2 removal efficiency, may corrode FGD constituents, and may require the purge of additional waste streams. As another example, describe the effects of different coals on the resultant mixture of combustion gases (e.g., percent oxygen in exhaust) and the resulting effects on the chemical reactions and efficiency of the scrubber. Define the scrubber system parameters which are controlled to compensate for these variations.

(4) Describe the purpose, need, and methods for reheat of exhaust gases downstream of FGD systems. Define energy consumption for reheat of these gases for typical plants ranging in size from 25 to 1000 megawatts for the different FGD systems. Describe problems experienced with reheaters and design and operating or maintenance factors used to satisfactorily overcome these problems. Assess approaches which have been or could be used to eliminate reheat of these gases.

As used herein, the term "SO_2 removal efficiency" is limited to performance obtained from a single module of removal equipment over a short time, and tested over a period of several weeks to show that scale accumulation is not likely. The term does not reflect the longer periods of time necessary for commercial operations, nor any implication concerning the equipment reliability or availability. Inasmuch as increasing the percent SO_2 removal above currently demonstrated levels will require greater chemical work, it will inherently cause some reduction in the equipment reliability and availability unless suitable counteracting steps (such as the use of more modules) are taken. It will also somewhat increase the energy consumption requirements, the chemical feed requirements, and the waste disposal requirements over those currently being used. It is understood that these specific subjects, their quantification, and potential solutions are being addressed by others and that their results will be integrated with the results reported herein. Thus, the results contained in this report should not be construed as a complete assessment of all factors related to the judgment as to the establishment of SO_2 removal New Source Performance Standards.

ABSTRACT

A brief introduction is given to flue gas desulfurization technology. The SO_2 removal mechanisms for wet scrubbing and dry processes are described, along with the typical gas contacting equipment used. Processes and their critical operating parameters are described for:

- Lime and limestone scrubbing
- Sodium carbonate scrubbing
- Double alkali (soda lime) scrubbing
- Magnesium oxide scrubbing
- Sodium sulfite (Wellman Lord) scrubbing

The report describes the implications of requiring 90 percent or greater SO_2 removal for new FGD installations. For lime and limestone scrubbing data are presented from the EPA Shawnee Test Facility to show the effects of process operating parameters on SO_2 removal efficiency. Mathematical correlations of the data are also presented. Additional test data is presented for sodium carbonate and double alkali scrubbing.

Coal characteristics and their effects on FGD system design and operation are summarized. A qualitative description is given for the effects of coal heating value and sulfur, ash, moisture, and chlorine content.

The reasons and methods for reheating scrubbed flue gases are described. Energy consumption of the various methods is defined, and problems experienced with reheater design, operation, and maintenance are presented. Alternative approaches to eliminating reheat are discussed.

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Section 1

SUMMARY

FGD PROCESSES

All flue gas desulfurization (FGD) installations operating on U.S. coal-fired utility boilers are wet scrubbing processes that contact the flue gas with a water solution of soluble (or sparingly soluble) alkali to absorb the SO_2. The SO_2 removal is carried out in devices (scrubbers) designed to provide intimate contact between gas and liquid over a large gas-liquid interface area. The SO_2 removal efficiency of an FGD system is a function of the type of scrubber selected, the absorbent composition and circulation rate to the scrubber, and additional parameters.

Of the potentially available processes, lime and limestone slurry scrubbing have received the most attention and are presently operating on a number of coal-fired boilers. Additional new units are under construction. These processes achieve high SO_2 removal efficiencies, use naturally occurring raw materials, produce relatively insoluble wastes, and have lower capital and operating costs than many other processes. However, they require careful design to prevent problems such as scaling and plugging in the scrubbers and erosion and corrosion of equipment.

Sodium carbonate scrubbing is a clear liquor absorbent process that is less susceptible to scaling than calcium-based systems. This process has had limited application because it consumes an expensive alkali and produces a soluble sodium waste product that must either be

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regenerated at a substantial cost or impounded in suitably lined retention ponds. To date, application has been limited to geographical areas having a cheap source of low grade carbonate; these include three units operating at a power station in Nevada and one under construction in Wyoming. Both stations burn low sulfur coal and the FGD systems are designed to realize about 90 percent removal efficiency.

The double alkali process retains the desirable characteristics of the clear liquor sodium carbonate system. It minimizes the soluble waste problem by reacting the scrubber effluent with lime or limestone to precipitate an insoluble waste product and regenerate the costly sodium absorbent. The process has achieved high SO_2 removal efficiencies on industrial boilers and greater than 90 percent removal during prototype testing on a coal-fired utility boiler. The first full-scale coal-fired utility application is scheduled to start up in 1979. The system is designed to meet performance guarantees of 95 percent removal on 5 percent sulfur coal. Construction of two additional units is planned. The process operation is somewhat more complex than lime or limestone scrubbing.

The magnesium oxide slurry scrubbing process features recovery of sulfuric acid or sulfur. It has been demonstrated to be feasible on full-scale coal-fired boilers and is capable of SO_2 removal efficiencies greater than 90 percent, although the reliability of these installations must be improved. A 120 MW unit is currently undergoing testing and has been designed to achieve 90 percent SO_2 removal from 2.5 percent sulfur coal. No additional systems are under construction.

The Wellman Lord process recovers either sulfuric acid or sulfur and has demonstrated SO_2 removal efficiencies greater than 90 percent on oil-fired boilers and industrial plants. The first coal-fired boiler application, a 115 MW installation, is currently undergoing a comprehensive one-year demonstration program. The unit was designed to meet performance guarantees of 90 percent SO_2 removal from 3.5 percent sulfur coal and has demonstrated greater efficiencies. Two additional Wellman Lord installations are under construction. The process operation is more complex and costly than lime and limestone scrubbing.

A number of FGD system parameters must be monitored and controlled to ensure proper operation, to accommodate variations such as changing boiler load and inlet SO_2 conditions, to ensure proper mist eliminator operation, and to prevent scaling in scrubbers. These parameters include a varying alkali feed rate to match inlet SO_2 and boiler load changes, and a number of mist eliminator washing schemes and design configurations to maintain trouble-free operation. Techniques employed to prevent scaling typically include control of scrubber pH, adequate hold tank residence time, control of suspended solids concentration in slurry systems, and high liquid-to-gas ratios. Clear liquor scrubbing processes such as the sodium carbonate, double alkali and Wellman Lord are not susceptible to scaling problems to the extent that the lime and limestone processes are.

IMPLICATIONS OF REQUIRING 90 PERCENT OR GREATER SO_2 REMOVAL FOR NEW FGD INSTALLATIONS

For the FGD processes described in this report, SO_2 removal efficiencies of 90 percent or greater — as defined in the foreword to this report — are obtainable by suitably changing a number of the operating conditions. For lime and limestone scrubbing, the process variables requiring change may include scrubber type, liquid-to-gas ratio, gas velocity, scrubber inlet pH, additives such as magnesium, and the number of scrubbing stages.

An FGD system is designed to achieve a specified SO_2 removal efficiency for the most severe operating conditions anticipated, such as maximum boiler load (equivalent to maximum gas volume to be treated) and maximum sulfur content of the coal to be burned (equivalent to maximum SO_2 concentration in the gas). The removal efficiency is largely fixed by the hardware and by the chosen level of operating parameters. Once these parameters are set, the system should achieve the design SO_2 removal at maximum boiler load and SO_2 concentration. For lower inlet SO_2 concentration, performance is generally improved for lime and limestone scrubbing. For some types of scrubbers, such as spray towers, efficiency may be improved with a reduction in boiler load. For others, such as tray towers, performance may be impaired and the FGD system must be carefully designed to achieve the required removal efficiency during boiler turndown.

For a given coal and FGD system, increasing the sulfur removal increases the sulfur load on the system, and with it, the associated cost, complexity, and energy consumption. On the other hand, for lime and limestone scrubbing a higher SO_2 removal efficiency can normally be achieved when using a low-sulfur coal, than when using a high-sulfur coal, with substantially less complexity. The ease of achieving a specified SO_2 removal efficiency depends not only upon the system or apparatus employed, but upon the coal sulfur content and heating value as well.

Raising the design efficiency of a lime or limestone FGD system from 60 or 85 percent to 90 percent or greater affects essentially all process design parameters and equipment specifications. Problems related to nonuniform gas and liquid distribution and scrubber turndown capabilities are amplified as the design efficiency is increased. Bypass reheat would be marginal or impossible for 90 percent SO_2 removal because all the flue gas must be treated.

EFFECT OF COAL PROPERTIES

The major coal properties affecting FGD system design and operation are heating value, sulfur, ash, moisture, and chlorine content.

Lower heating value coals generally give rise to greater flue gas volumes and SO_2 emissions per megawatt of generated power. On a megawatt basis, the FGD system is larger and capital and operating costs are higher for coal with lower heating value (for a given sulfur content and absorption efficiency). A characteristic of lower heating value coals and coals of high moisture content is a flue gas with a greater proportion of water vapor. This leads to smaller amounts of water evaporation in the scrubbers, which in turn leads to higher gas and recirculating liquor temperatures. Elevated temperatures could affect absorption efficiency, depending on the chemistry of the particular process.

The sulfur content together with the allowable emission standards determines the required SO_2 absorption rate. For a given absorption efficiency, the sulfur content affects the entire FGD system design. The high SO_2 removal efficiency required for high sulfur coals leads to a more complex and costly FGD system. Low sulfur coals tend to promote higher sulfite oxidation, which may be desirable or undesirable depending on the process.

FGD processes using nonregenerable absorbents (lime, limestone, or soda) can be used for combined removal of fly ash and SO_2. The fly ash is then disposed of together with the spent absorbent. However, it may still be desirable to remove the fly ash upstream of the FGD system since it can erode process equipment, interfere with process chemistry (and cause scaling in the soda system), and contribute to the waste volume. Some fly ash from low sulfur western coals has a high alkali content that can be used in the absorption process.

The small amounts of chlorine in coal are converted to gaseous chloride in the boiler. The chloride is absorbed from the gas by wet scrubbing processes. Its presence provides the potential for chloride stress-corrosion, requiring the use of high alloy equipment wherever rubber or other protective coatings are not applicable. Alkali associated with the chloride is lost as dissolved solids in the waste stream and, because of its solubility, the waste may require special disposal. Chloride may also affect system chemistry. Prescrubbing may be used to absorb chlorides from the flue gas upstream of the FGD system.

REHEAT OF SCRUBBED FLUE GAS

Flue gas normally enters a wet FGD system at a temperature between 120 and 150°C (250 and 300°F). When contacted with the scrubbing liquor, the gas is cooled by evaporation to the water dew point of about 50°C (125°F). Discharge of cool, wet gas to the stack can lead to: condensation and the resulting acidic water corrosion and fly ash fouling of ductwork, fans, and stacks; acid rain from the plume; impaired plume rise and dispersion; and increased plume opacity.

Reheating the flue gas 15 to 30°C (about 30 to 50°F) or more is frequently required to eliminate the above problems. Methods of reheating include: steam or pressurized hot water heated tubular exchangers in direct contact with the flue gas; steam heated finned exchangers used to heat air for blending with the flue gas; blending hot combustion gases from an oil or gas burner with the flue gas stream; and bypassing a portion of the hotter untreated gas around the FGD system for mixing with the scrubbed gas.

Reheat is undesirable because it consumes sizable amounts of energy. Alternatives to reheat include tall stacks to enhance plume dispersion, corrosion protection with high alloy steels or noncorrosive coatings and linings, upstream warm gas fans, and efficient mist eliminators. A number of FGD installations operate without reheat under "wet-stack" conditions. Thus, reheat does not seem to be a universal requirement for wet scrubbers.

SECTION 2

FGD PROCESSES

FGD TECHNOLOGY SUMMARY

More than 50 processes are in various phases of development and commercial application for the removal of SO_2 from industrial waste gases and boiler flue gases. The processes differ in the manner in which they remove SO_2 from the gas and in the physical form in which the absorbed sulfur is discharged. Important classifications of flue gas desulfurization (FGD) processes that have been tested on coal-fired boilers are shown in Figure 2-1.

A major distinction is generally made between recovery processes, which recover SO_2 in useful forms such as sulfuric acid or elemental sulfur, and throwaway processes, which produce a solid or liquid waste. In the throwaway case, the waste product must be stored in a pond or treated and used for landfill.

Throwaway and recovery processes have been developed that operate in either wet, dry, or semidry modes. Wet scrubbing processes contact the flue gas with a large and generally recirculated flow of a water solution or slurry. The scrubbing liquor absorbs the SO_2 and cools the gas by evaporation of water to a temperature only slightly above the water dew point. It may then be desirable to reheat the gas to minimize condensation, fouling, and corrosion in ducts, fans, and stacks; to avoid a visible plume; and to improve plume rise and dispersion. Dry processes have an advantage in not requiring stack gas reheat because the flue gas is not contacted by water. Use of a spray drier

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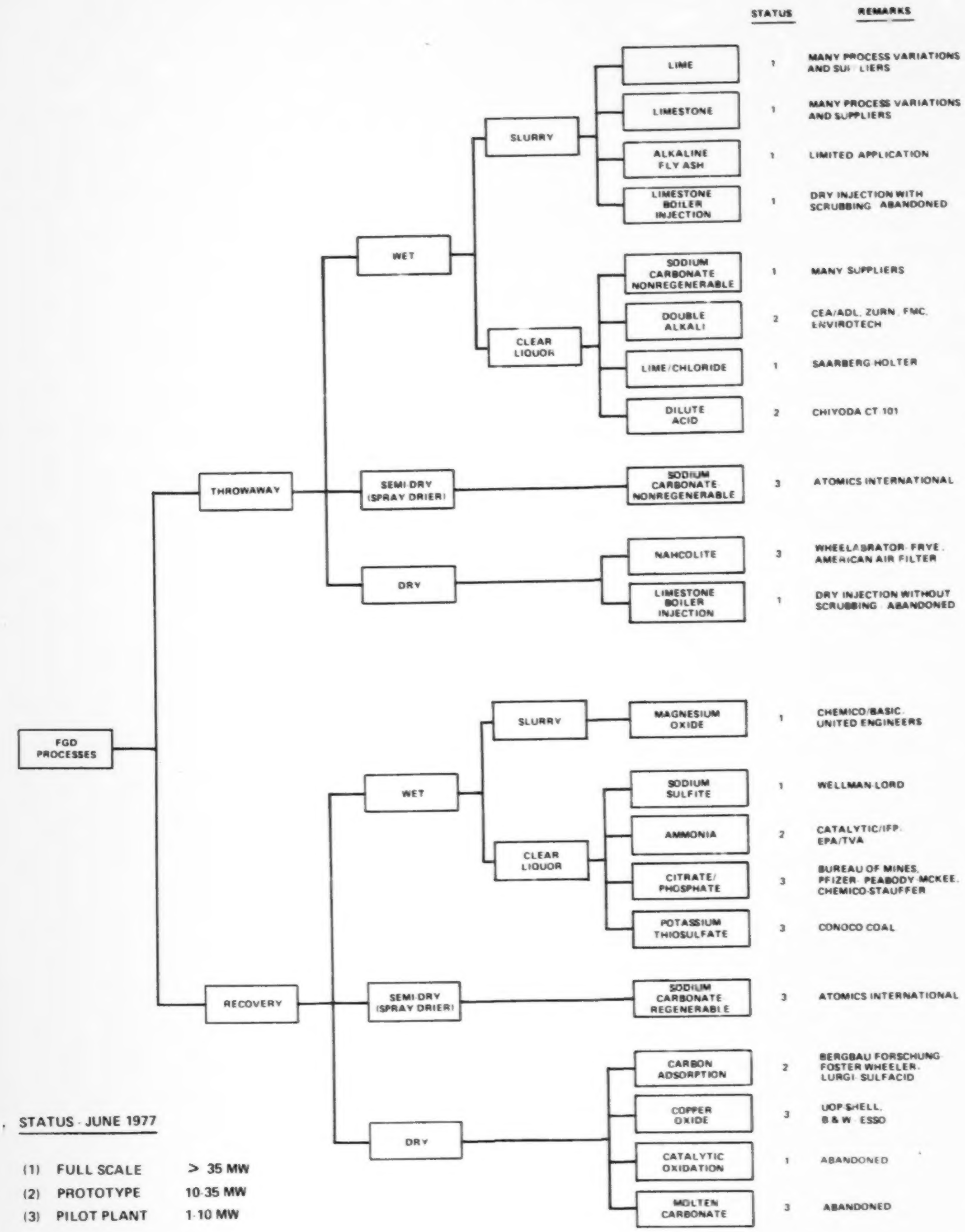
STATUS
REMARKS
FGD PROCESSES
THROWAWAY
WET
SLURRY
LIME
1
MANY PROCESS VARIATIONS AND SUPPLIERS
LIMESTONE
1
MANY PROCESS VARIATIONS AND SUPPLIERS
ALKALINE FLY ASH
1
LIMITED APPLICATION
LIMESTONE BOILER INJECTION
1
DRY INJECTION WITH SCRUBBING ABANDONED
CLEAR LIQUOR
SODIUM CARBONATE NONREGENERABLE
1
MANY SUPPLIERS
DOUBLE ALKALI
2
CEA/ADL, ZURN, FMC, ENVIROTECH
LIME/CHLORIDE
1
SAARBERG HOLTER
DILUTE ACID
2
CHIYODA CT 101
SEMI-DRY (SPRAY DRIER)
SODIUM CARBONATE NONREGENERABLE
3
ATOMICS INTERNATIONAL
DRY
NAHCOLITE
3
WHEELABRATOR-FRYE, AMERICAN AIR FILTER
LIMESTONE BOILER INJECTION
1
DRY INJECTION WITHOUT SCRUBBING ABANDONED
RECOVERY
WET
SLURRY
MAGNESIUM OXIDE
1
CHEMICO/BASIC, UNITED ENGINEERS
CLEAR LIQUOR
SODIUM SULFITE
1
WELLMAN-LORD
AMMONIA
2
CATALYTIC/IFP, EPA/TVA
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BUREAU OF MINES, PFIZER-PEABODY-MCKEE, CHEMICO-STAUFFER
POTASSIUM THIOSULFATE
3
CONOCO COAL
SEMI-DRY (SPRAY DRIER)
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DRY
CARBON ADSORPTION
2
BERGBAU FORSCHUNG, FOSTER WHEELER, LURGI-SULFACID
COPPER OXIDE
3
UOP-SHELL, B & W-ESSO
CATALYTIC OXIDATION
1
ABANDONED
MOLTEN CARBONATE
3
ABANDONED
STATUS - JUNE 1977
(1) FULL SCALE > 35 MW
(2) PROTOTYPE 10-35 MW
(3) PILOT PLANT 1-10 MW

Figure 2-1. Summary of FGD Processes Tested on Coal-Fired Boilers

results in a semidry mode of operation in which the gas is contacted by small quantities of aqueous absorbent. In this case, the absorbent is dried, the flue gas is only partially cooled, and reheat can be avoided.

Wet Processes

Some wet scrubbers are suitable for the removal of both fly ash and SO_2 so that, for a new power plant, one of the options for the control of these pollutants may be to let the FGD system remove both. In practice, however, there are some valid reasons for collecting fly ash separately with an electrostatic precipitator. For example, possible interference with process chemistry is prevented, erosion of process equipment by fly ash slurry is minimized, contribution to additional throwaway sludge volume is avoided (although the fly ash must still be disposed of), and, in the case of recovery processes, contamination of by-products is prevented. Another consideration is that if the power plant has an FGD system bypass and separate dust removal equipment, particulate removal operation can be continued during periods of FGD system maintenance (if such operation is allowed).

Wet FGD processes fall into two classifications - slurry scrubbing and clear liquor. In slurry scrubbing the reactive absorbent and the absorbent-SO_2 reaction products are present as suspended solid particles in water. During the absorption phase of slurry scrubbing, the suspended alkali dissolves to react with the absorbed SO_2 and to precipitate waste solids which are disposed of in throwaway processes or treated to regenerate active alkali in recovery processes. The circulation of saturated solutions and crystallization of reaction products makes the control of process chemistry very important to prevent scaling in the scrubber. Slurry system components are also vulnerable to plugging and erosion. Clear liquor scrubbing processes minimize the problems of the slurry systems by keeping the absorbent and reaction products dissolved in water.

Dry Processes

Dry FGD processes, in most cases, operate at low gas velocities, thereby requiring large vessels, while wet systems operate at higher velocities with smaller vessels. The dry carbon adsorption processes treat flue gases at temperatures lower than 150°C (300°F), which means that they can be located downstream from the power plant air heater and, therefore, can be added to existing power stations. The copper-oxide and catalytic-oxidation processes require flue gas temperatures above 370°C (700°F). Therefore, they would require reheat as an add-on system, and their full potential would be better realized by integrating them upstream of the air heater in a new installation. Nahcolite injection can be carried out at temperatures above or below 150°C (300°F), although better efficiency is realized at higher temperatures.

Dry and Semidry Processes

The dry and semidry throwaway processes produce a dry waste product that may be more easily disposed of than the sludge or moist waste solids from wet throwaway processes.

Commercial Processes on U.S. Boilers

Table 2-1 summarizes by process the number and capacity of full scale operational FGD systems and systems under construction on utility boilers in the U.S. The lime and limestone processes have received the most attention. These throwaway processes achieve high SO_2 removal efficiencies and have capital and operating costs that are lower than many other processes. On the other hand, they require careful design to minimize or prevent problems with erosion, corrosion, sludge disposal, and scaling and plugging in the scrubbers.

Alkaline fly ash scrubbing is of interest because it utilizes the alkaline values in the fly ash of some western low sulfur coals, thereby minimizing raw material requirements. Its application is, therefore, limited to coals with the appropriate ash qualities.

Table 2-1

SUMMARY OF FGD PROCESSES ON U.S. COAL-FIRED UTILITY BOILERS

Process	Operational		Construction		Contract Awarded	
	No.	MW	No.	MW	No.	MW
Limestone	9	3,767	16	6,750	10	4,911
Lime	9	2,702	9	4,158	4	2,350
Alkaline fly ash with lime or limestone	2	720	1	450	2	1,400
Sodium carbonate	3	375	–	–	1	509
Magnesium oxide	1	120	–	–	–	–
Double alkali	–	–	1	575	2	527
Wellman Lord	1	115	2	715	–	–
Aqueous carbonate	–	–	–	–	1	100

Reference 1

June 1977

Sodium carbonate scrubbing is a clear liquor absorbent process which is less susceptible to scaling than calcium-based systems. This process has had limited application because it consumes an expensive alkali and produces a soluble sodium waste product which must either be regenerated at substantial cost or impounded in suitably lined retention ponds.

The double alkali process retains the desirable characteristics of the clear liquor sodium carbonate system, but avoids the soluble waste problem by reacting the scrubber effluent with lime or limestone to precipitate an insoluble waste product and regenerate the costly sodium absorbent. However, the process is more complex and requires additional equipment. It has not yet been tested in full-scale operation on a coal-fired boiler.

The magnesium oxide recovery process is less prone to scaling than lime or limestone processes and features recovery of sulfuric acid or sulfur. It has been demonstrated to be feasible on full-scale coal-fired boilers, but the reliability of these installations must be improved.

The Wellman Lord system has been the most widely installed recovery process for oil-fired boilers in Japan. It recovers either sulfuric acid or sulfur and is currently undergoing full-scale demonstration on a high-sulfur coal-fired boiler.

Modifications to existing processes and new processes with features designed to overcome the drawbacks of current systems will continue to be developed. In the U.S. full-scale coal-fired demonstration units are planned for the semidry sodium carbonate recovery process and for the clear liquor citrate recovery process. In Germany the Saarberg-Hölter clear liquor lime/chloride scrubbing process has been operating successfully over 14,000 hours on a coal-fired boiler (Ref.2).

SO_2 REMOVAL MECHANISMS

Removal of SO_2 from flue gas can be accomplished by absorption in liquids and by sorption mechanisms on solids. In absorption, the SO_2 passes through the gas-liquid phase boundary to become distributed throughout the liquid and undergo chemical reaction with the absorbent. With the use of solid adsorbents, the SO_2 may be retained on the surface by mechanisms that are either physical or chemical in nature.

Wet Processes

In the absorption of SO_2 by a liquid, the SO_2 must diffuse out of the gas phase and into the liquid phase. The driving force for the transfer of SO_2 from the gas to liquid phase is provided by the concentration differences between the SO_2 in the flue gas and the equilibrium vapor pressure, or back pressure, of the absorbed SO_2 in the liquid. The SO_2, when absorbed in the liquid, has a finite back pressure that increases in proportion to its concentration in the liquid. When this back pressure equals the SO_2 concentration in the flue gas, no further absorption can take place.

In the selection of an absorbent for scrubbing flue gas, the object is to find a liquid with capacity to absorb a large quantity of SO_2 without building up an appreciable equilibrium back pressure. This is accomplished by using a solution with high soluble alkali content; e.g., lime, limestone, sodium carbonate, sodium sulfite, magnesium sulfite. The wet processes are designed to operate so that the SO_2 back pressure over the incoming absorbent is essentially zero (less than 1 ppm SO_2).

With clear liquor scrubbing, the absorbent and chemical reaction products of absorbed SO_2 with absorbent remain dissolved in water. In slurry systems such as lime and limestone, however, the absorbents and reaction products are of limited solubility, so that both are present as suspended solids in the scrubbing slurry. During the absorption step of slurry scrubbing, the suspended alkali dissolves in the scrubber to react with the absorbed SO_2 and precipitate the reaction products.

Design Parameters. The SO_2 absorber has as its objective the intimate contact between gas and liquid over a large gas-liquid interfacial surface area. Important design parameters in the selection of a scrubber are:

- Liquid-to-Gas Ratio. The liquid-to-gas ratio (L/G), expressed as liters of liquid pumped through the scrubber per cubic meter of gas passing through the scrubber (l/m^3), or gallons per thousand cubic feet (gal/Mcf), is related to the gas-liquid interfacial area in the scrubber and the quantity of absorbent available for reaction with the SO_2. The surface area may be generated largely by production of liquid spray through nozzles or by atomization by action of the gas flow. The liquid-to-gas ratio strongly affects the system energy losses. High L/G ratios introduce major piping and structural design considerations which are reflected in higher capital costs as well

- Gas Velocity. To minimize equipment cost, scrubbers are designed to operate at maximum practicable gas velocities, thereby minimizing vessel size. Maximum velocities are dictated by gas-liquid distribution characteristics and by the maximum allowable liquid entrainment that the mist eliminator can handle. Gas velocities may be 1.5 to 10 m/sec (5 to 30 ft/sec) in tower-type scrubbers and more than 30 m/sec (100 ft/sec) in the throat of a venturi scrubber. The lower the velocity, the less the entrainment, but the more costly the scrubber

- Gas Turbulence. To promote rapid diffusion of SO_2 from the bulk gas to the liquid interface, a high degree of turbulence is desirable. With a clear liquor absorbent that is not limited by slow dissolution of suspended alkali particles, gas residence times of a few hundredths of a second are adequate to obtain high SO_2 absorption efficiencies such as in a venturi scrubber

- Gas Pressure Drop. The energy required to move the flue gas through the scrubber is measured in terms of gas pressure drop through the scrubber, expressed as millimeters of mercury (mm Hg), or inches of water (in. H_2O)

- Slurry Holdup. For slurry scrubbing processes, large liquid holdup, or residence time, is desirable to provide time for dissolution of alkali so that a maximum amount is utilized through the scrubber. Residence times in towers with packing may be 5 seconds or more. Residence times in venturi scrubbers are a few hundredths of a second, too small for high absorption efficiencies with lime or limestone slurry scrubbing (unless additives are provided — see Section 4)

- Gas Distribution. A major problem in the scaleup of scrubbers from pilot to commercial size has been maintaining uniform gas distribution. If the flow is not uniform the scrubber will not operate at design SO_2 absorption efficiencies. In practice, uniform flow has been difficult to achieve. Typically, turning vanes near the scrubber inlet duct and compartmentalization have been employed.

- Scrubber Internals. To promote maximum gas-liquid surface area and liquor holdup a number of devices are used. Common methods include stages of moving spheres supported on grids, towers filled with packing, towers with closely spaced rods or grids, and perforated trays. Fixed packings are generally not used in slurry service because of a tendency to plug or, in lime or limestone service, to scale. The absorbent is often sprayed into the towers through several stages of spray headers

- Turndown. To follow changes in boiler load it is necessary that the scrubber provide good gas-liquid distribution, high liquid holdup for some processes, and high gas-liquid interfacial area over varying gas flow rates. Some scrubbers can be turned down to 50 percent of design, while others must be divided into sections that can be closed off. A venturi can have a variable throat area to accommodate turndown. In a big installation, individual modules can be taken out of service

- Gas-Liquid Flow Configuration. Scrubbers can be counter-current (the gas flows upward and the liquor downward), co-current (gas and liquor flow downward together), or cross-current (gas flows generally horizontally and liquor is sprayed at a right angle to the flow). Theoretically, counter-current or cross-flow would be best because the scrubbed gas is being continually contacted by the fresher absorbent, but multiple spray stages in co-current flow can be designed to achieve the same effect. The gas flow configuration directly affects the mechanical layout and duct work

- Mist Elimination. A mist eliminator is installed in the scrubber or downstream of the scrubber to separate mist carried over in the cleaned gas. This is done primarily to limit the emission of particulates to the atmosphere; to prevent corrosion of downstream ductwork, fans, and stack lining; to prevent plugging of inline reheater tubes and solids deposition on fan blades which could cause imbalance; and to reduce reheater energy requirements by removing water mist from the flue gas stream

Wet Scrubbers. Leading scrubber types include the following:

- Venturi. The venturi scrubber (Figure A-1) has been used when both fly ash and sulfur dioxide must be removed. The venturi has been able to remove fly ash down to 0.05 gm/m^3 (0.02 gr/scf) with pressure drops of 20 to 30 mm Hg (10 to 15 inches H_2O) and liquid-to-gas ratios of 1.5-4.0 l/m^3 (10 to 30 gal/Mcf) for typical dust loadings and particle size distributions from power plant stack gases. The venturi may contain an adjustable throat area to permit control of pressure drop over a wide range of flow conditions. The SO_2 absorption efficiency is limited to 30 to 50 percent per stage with lime-limestone because of the short liquid residence time. It therefore requires additives (such as MgO) or two stages of venturis or an after-absorber to achieve higher absorption efficiencies than this with high sulfur fuels

- Turbulent Contact Absorber (TCA). The TCA (Figure A-2) is a counter-current multistage scrubber with screens that both support and restrain the plastic spheres. The spheres move in a turbulent fashion providing good gas-liquid contact and scale removal. With additive-free calcium systems, the number of stages are generally between two and four for maximum SO_2 absorption efficiency with liquid-to-gas ratios of 6 to 8 l/m^3 (40 to 60 gal/Mcf). The pressure drop per stage is approximately 4 to 5 mm Hg (2 to 2.5 inches H_2O)

- Spray Tower. The spray tower (Figure A-1) is a counter-current type scrubber with a low gas pressure drop. Absorbent is sprayed through several headers, and liquid-to-gas ratios of 10 l/m^3 (80 gal/Mcf) or more are required for maximum SO_2 absorption with additive-free calcium systems

- Packed Tower. The packed tower is a very efficient type of scrubber, but has seen little use with slurry scrubbing because of susceptibility to plugging. With clear liquor scrubbing, high absorption efficiencies may be achieved with liquid-to-gas ratios of 1.5 to 4 l/m^3 (10 to 30 gal/Mcf)

- Marble-Bed Absorber. The marble-bed absorber uses a 10-cm (4-inch) bed packing of glass spheres (marbles) that are in slight vibratory motion. A turbulent layer of liquid and gas above the glass spheres increases mass transfer and particulate removal. Pressure drop is generally 7.5 to 11 mm Hg (4 to 6 inches H_2O). Liquid-to-gas ratios of 3 to 4 l/m^3 (25 to 30 gal/Mcf) have been used with additive-free calcium systems. Use of this device is no longer advocated by its foremost developer because of scaling and distribution problems encountered over a number of years

- Tray Column. The tray column offers high liquid holdup and absorption efficiency at relatively low pressure drop. The main disadvantage is its vulnerability to scaling. Liquid-to-gas ratios of 5 l/m^3 (40 gal/Mcf) are required for additive-free calcium systems. In addition, undersprays are used to wash off soft scale deposits, consuming further energy

- Cross-Flow Absorber. The cross-flow absorber is installed in a horizontal position. It has a low pressure drop and has been tested with packing or sprays. It requires a high L/G ratio to realize high absorption efficiency with additive-free calcium systems

- Screen or Grid Tower. The screen or grid scrubber contains five to ten screens or grids (22 mm, or 7/8-inch openings). A liquid-to-gas ratio of 7 or more l/m^3 (50 gal/mcf) is used with additive-free calcium systems. Pressure drops are lower for this type of absorber than for tray or packed tower types

- Venturi Rod Absorber. The venturi rod absorber utilizes closely spaced parallel circular rods placed perpendicular to the gas flow in the absorber. The circular cross-section of the rods provides a converging-diverging gas flow path (similar to a venturi). This type of scrubber is marketed by Riley Stoker under the trade name Ventri Rod.

Energy Requirements. Wet scrubbing systems have significant energy requirements to overcome fan and pumping losses. Although there can be considerable variation, Figures 2-2 and 2-3 (Ref. 3) show the order-of-magnitude expected for typical scrubbers. Other energy losses include discharge pumps, small auxiliaries, lighting, instrumentation, and additional process equipment, and also thermal energy consumed by gas reheat (where used) and associated fan power to overcome added pressure losses.

Wet Scrubber Efficiency. For a given process, the design efficiency is determined by knowledge of the system characteristics. Selection of scrubber type can provide high or low liquid pumping or gas pressure drop energy input, contacting area for the phases, and scrubber residence time. For example, a spray tower has low gas pressure drop and moderate contact area and residence time, while a venturi has high pressure drop and contact area, but very low residence time. In a packed tower, increased packing (at the expense of higher pressure drop) tends to increase SO_2 removal because of the increased gas-liquid interfacial area provided. High L/G provides more liquid surface area for absorption, and also provides a larger capacity for absorption. Increased gas velocity can be detrimental to removal because of decreased gas-phase residence time and lower L/G for fixed liquor pumping rate. However, these effects can be offset by the higher energy input, which increases agitation and contact surface area for some types of scrubbers (TCA and venturi scrubber, but not spray tower). The design efficiency represents a trade-off between maximizing removal and minimizing cost and, most importantly, must be consistent with demonstrated high availability as well.

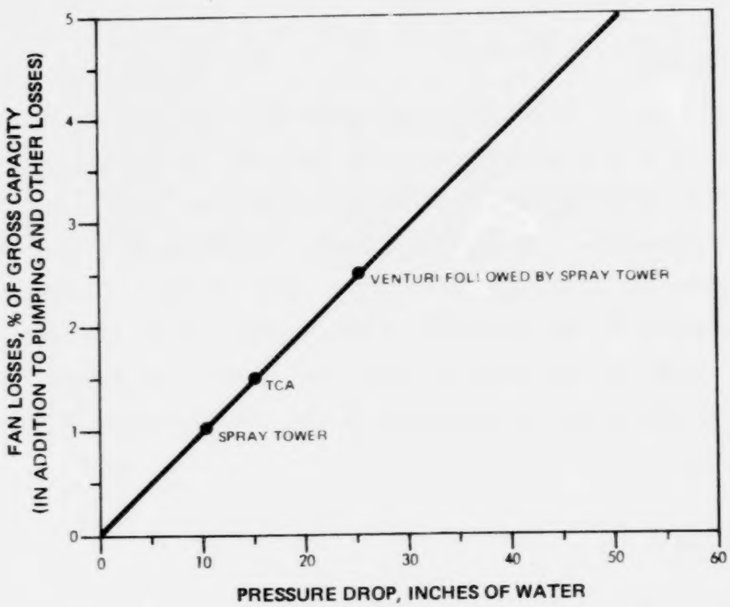

Figure 2-2. Station Electrical Loss as a Function of Draft Requirements

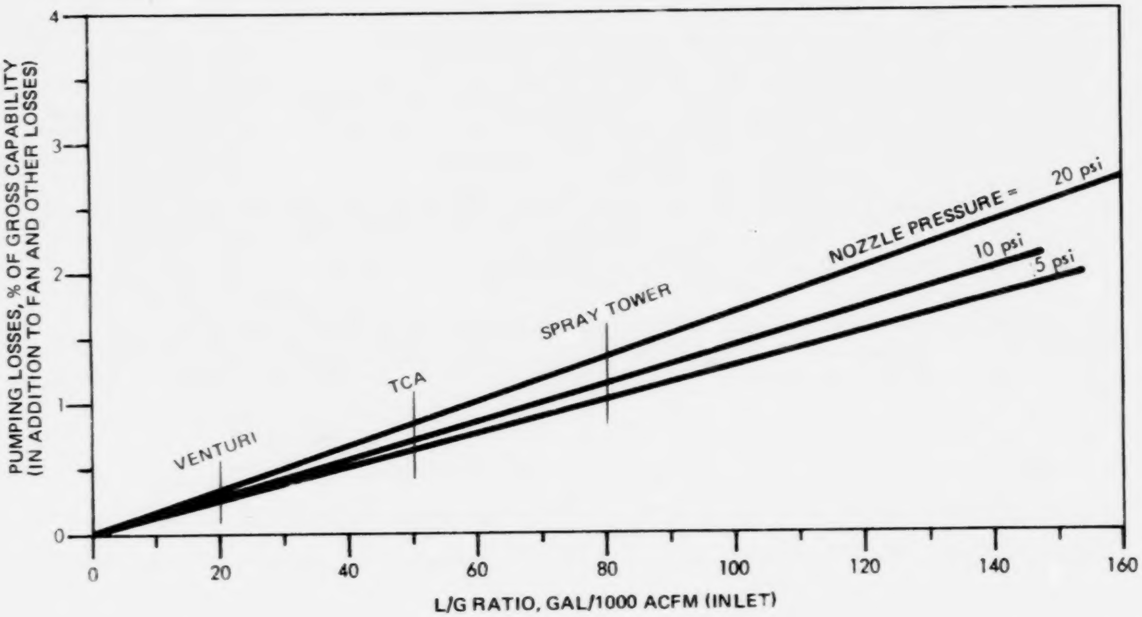

Figure 2-3. Station Electrical Loss as a Function L/G Ratio and Nozzle Pressure

Semidry Process

The use of a spray drier has been proposed (Ref. 4) for SO_2 liquid absorption as a process whereby the flue gas is contacted by very small quantities of sprayed sodium carbonate solution (L/G less than 0.13 l/m^3, or 1 gal/Mcf). In a spray drier, droplets generated by a centrifugal atomizer are much finer than, for example, those in a spray tower. Because of the low L/G, there is not enough water to saturate the flue gas. Only slight cooling takes place so that reheat is not needed. Because all the water is evaporated, the spent reactants can be removed as dry particles.

Dry Processes

The primary feature of dry FGD processes is removal of SO_2 without saturating the flue gas or significantly lowering its temperature. Dry processes involve several different mechanisms, including:

- SO_2-solid reaction with an alkali or metal oxide dispersed on a high surface area carrier
- Chemisorption with oxygen and water on the surface of a catalytic solid

The ultimate disposal of the SO_2 from a dry system may be by:

- Discarding the reaction products
- Regenerating the SO_2 acceptor solid by liberating a concentrated stream of SO_2 for subsequent recovery or processing

Nahcolite. Short-term pilot plant tests have been performed (Ref. 5) in which a dry powdered alkali (sodium bicarbonate or nahcolite) was injected directly into a slip stream of boiler flue gas. In this case the SO_2 reacts directly with the nahcolite to form a dry sodium sulfate product which can then be collected in a baghouse, or other particulate collector, along with the fly ash. The process is somewhat more efficient at higher temperatures (290 to 320°C, or 550 to 600°F), and up to 90 percent SO_2 removal has been claimed.

Activated Carbon Adsorption. A number of FGD processes employ the concept of adsorption of the SO_2 onto the surface of a char or activated carbon. In this process, oxygen and water vapor are also adsorbed onto the char, which catalyzes formation of sulfuric acid from the SO_2, oxygen, and water vapor. The acid is adsorbed on the carbon surface. When the charcoal bed becomes loaded with SO_2, it must be regenerated for further use. Suggested methods for regeneration include:

- Thermal Regeneration. The SO_2 loaded char is heated to about 650°C (1,200°F) at which temperature the adsorbed acid in the pores reacts with the carbon and is reduced to SO_2. A gas stream of concentrated SO_2, CO_2, and water is exhausted from the generator and further treated to produce sulfur or sulfuric acid. This is the method used in the Bergbau Forschung-Foster Wheeler process (Ref.6)

- Water Wash Regeneration. The charcoal bed is sprayed with water, which removes the sulfuric acid formed in the carbon pores. The dilute sulfuric acid is collected and disposed of. This is the method piloted in the Lurgi Sulfacid process and used in the Hitachi process, which to date has been limited to oil-fired boilers (Ref. 7)

- Chemical Regeneration. In this method the loaded carbon bed is brought into contact with hydrogen sulfide to produce elemental sulfur. This is the Westvaco process concept (Ref. 8)

The heart of the typical SO_2 adsorption process is the gas-solid contactor. The objective is to maximize contact of gas and adsorbent and minimize gas pressure drop, vessel size, and mechanical stress on the adsorbent. The two common types of contactors are:

- Moving-Bed Contactor. Charcoal is cycled continuously through the adsorption vessel (usually counter-current) and then through the regeneration vessel. The process is continuous. This is typified by the Bergau Forschung-Foster Wheeler method

- Fixed-Bed Contactor. The charcoal remains in place. The bed receives gas until loaded with SO_2, after which the flow is switched to a parallel contactor. The idle charcoal bed is then regenerated and adsorbed SO_2 removed for further processing

A drawback of all carbon adsorption processes is the necessary low gas velocity (0.3 to 0.6 m/sec, or 1 to 2 ft/sec) and, consequently, the large vessels or vessel-ductwork combination required.

Copper Oxide. In the copper oxide process (Ref. 9), the flue gas is blown through open passages parallel to a fixed bed of copper oxide acceptor material at 370 to 430°C (700 to 800°F). The SO_2 is removed from the gas to form copper sulfate. When the acceptor is fully sulfated, the gas is switched to a parallel contacting vessel, and the idle bed is contacted with a gas containing hydrogen. A concentrated stream of SO_2 is released for further processing. The copper sulfate is reduced to copper and then reoxidized during the start of the next adsorption cycle. This process has been applied to flue gas from refinery heaters in Japan, but these are only small test installations.

Catalytic Oxidation. In the catalytic oxidation process, the flue gas passes through a catalyst bed where the SO_2 is oxidized to SO_3. The SO_3 reacts with water vapor in the flue gas to form sulfuric acid. The gas is then cooled in a heat exchanger, and the sulfuric acid is condensed in a packed absorbing tower. The acid is collected and disposed of (or marketed). This process has been installed at the 100 MW level in the U.S. (Ref. 10), but has been unable to achieve sustained operation.

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Section 3

SELECTED FGD PROCESS DESCRIPTIONS AND CRITICAL SYSTEM OPERATING PARAMETERS

This section presents brief descriptions of the following flue gas desulfurization processes:

- Lime/limestone scrubbing
- Sodium carbonate scrubbing
- Double alkali (soda-lime) scrubbing
- Magnesium oxide scrubbing
- Sodium sulfite (Wellman Lord) scrubbing

Of these processes, lime, limestone, sodium carbonate and Wellman Lord systems are presently operating at full-scale coal-fired utility boilers, and additional new units are under construction. The magnesium oxide process is undergoing full-scale demonstration and a full-scale double alkali system is also under construction. The operational FGD systems on U.S. coal-fired utility boilers are summarized in Table 3-1.

LIME/LIMESTONE SCRUBBING

From the process standpoint, lime scrubbing and limestone scrubbing are similar operations that differ both with respect to feed slurry preparation and the internal chemistry. For a lime system, quick-lime, CaO, is first slaked in water to form calcium hydroxide, $Ca(OH)_2$. For a limestone system, the raw calcium carbonate stone, $CaCO_3$, is finely ground in a wet mill to form a makeup slurry.

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Process Description

A process flow diagram for a typical lime or limestone system is presented in Figure 3-1. Flue gas from the boiler enters the scrubber where it is contacted by a recirculating slurry containing the suspended alkali. The scrubbed gas leaves the absorber through a mist eliminator. While passing through the scrubber the gas is cooled and saturated with water vapor. The gas may have to be reheated to restore buoyancy to the plume and to prevent water condensation or solids deposition in the ductwork, fan, and stack (see Section 6).

During the absorption process the suspended alkali dissolves to react with the absorbed SO_2 and precipitate calcium sulfite and sulfate (gypsum) as solid waste. In Figure 3-1 the slurry from the scrubbers drains to a hold tank where reactions go to completion. Makeup lime is added to the tank and part of the slurry recycles back to the scrubbers. If fly ash is not collected upstream, it will be captured by the scrubber liquor. Typically, the continuously recycled scrubber slurry may contain from 5 to 15 percent suspended solids consisting of fresh alkali, reacted waste products, and fly ash.

To remove solid waste products from the system, a bleed stream is withdrawn from the recycle loop and routed to the solids separation equipment. In Figure 3-1, the bleed is routed to a clarifier (also called a thickener or settling tank). The suspended solids settle to the bottom of the tank, and clear liquor is drawn off the top and returned to the scrubber loop. At the bottom of the clarifier, the solids may be concentrated to 20 to 40 percent. This underflow then goes to a filter for further thickening, where the separated liquor is also returned to the scrubber loop and the filter cake, consisting of 60 percent or more solids, is discharged to a waste disposal area. The greater volume of clarifier underflow sludge may be routed directly

Table 3-1

OPERATIONAL FGD SYSTEMS ON U.S. COAL-FIRED UTILITY BOILERS

Plant Name and Company	Process	Vendor	New or Retrofit	Rating (MW)	% Sulfur in Coal	Fly Ash Collector	SO₂ Absorber	No. of Modules	L/G (gal/Mcf)	Design Efficiency %	Startup	Reference
Cholla 1 Arizona Public Service	Limestone	Research Cottrell	R	115	0.4-1	–	A-Venturi and packed tower B-Venturi	2	20 & 25 20	>90(a) 45	10/73	1,2
Duck Creek 1A Central Illinois Light	Limestone	Riley Stoker–Environeering	N	100	2.5-3	ESP	Ventri rod absorber	1	50	75-85	9/76(b)	1,3
La Cygne 1 Kansas City Power & Light	Limestone	Babcock & Wilcox	N	820	3.6	–	Venturi and sieve tray tower	7	33	60-80	2/73	1,3
Lawrence 4 Kansas Power & Light	Limestone(c)	Combustion Engineering	R	125	0.5	–	Marble bed–converted to venturi rod/spray tower in 1976	2	–	75-85	12/68	1
Lawrence 5 Kansas Power & Light	Limestone(c)	Combustion Engineering	N	400	0.5	–	Marble bed–being converted to venturi rod/spray tower in 1977	2	–	–	11/71	1
Martin Lake 1 Texas Utilities	Limestone	Research Cottrell	N	793	1.2 (lignite)	ESP	Spray quencher/wetted film contactor	6	80	97(d)	4/77	2
Sherburne 1 Northern States Power	Limestone	Combustion Engineering	N	710	0.8	–	Venturi rod/marble bed	12	27	50-60	3/76	1,3
Sherburne 2 Northern States Power	Limestone	Combustion Engineering	N	680	0.8	–	Venturi rod/marble bed	12	–	50-60	4/77	1
Southwest 1 Springfield City Utilities	Limestone	UOP	N	200	3.5	ESP	TCA	2	60	80-87	3/77	1,4
Widows Creek 8 TVA	Limestone	TVA	R	550	3.7	ESP	Venturi/spray tower	4	50	80	5/77	1
Will County 1 Commonwealth Edison	Limestone	Babcock & Wilcox	R	167	4	ESP	Venturi/two-stage perforated plate tower	2	34	85	2/72	3
Bruce Mansfield 1 Pennsylvania Power	Lime with MgO	Chemico	N	835	4.7	–	2-stage venturi	6	–	92	4/76	1
Cane Run 4 Louisville Gas & Electric	Lime	American Air Filter	R	178	3.5-4	ESP	Mobile bed	2	30-40	90	8/76	1,3
Conesville 5 Columbus & Southern Ohio Elec.	Lime with MgO	UOP	N	400	4.5-4.9	ESP	TCA	2	60	90	3/77	1,4

(a) Net removal is about 60% for two units.
(b) Shakedown operation only. Plant will burn low sulfur coal until 3 additional modules are built.
(c) Converted from boiler injection to conventional scrubbing–unit 4 in 1976, Unit 5 in 1977.
(d) Bypass reheat reduces net removal to about 60%.

Table 3-1 (Continued)

Plant Name and Company	Process	Vendor	New or Retrofit	Rating (MW)	% Sulfur in Coal	Fly Ash Collector	SO2 Absorber	No. of Modules	L/G (gal/mcf)	Design Efficiency %	Startup	Reference
Elrama Duquesne Light	Lime with MgO	Chemico	R	510	1-2.8	ESP	Venturi	5	30-35	85	10/75	3
Green River 1,2,3 Kentucky Utilities	Lime	American Air Filter	R	64	3.8-4	Mechanical	Venturi/mobile bed	1	35	>80	9/75	1,3
Hawthorn 3 Kansas City Power & Light	Lime(a)	Combustion Engineering	R	140	0.5-3.5	–	Marble bed	2	–	70	11/72	1
Hawthorn 4 Kansas City Power & Light	Lime(a)	Combustion Engineering	R	100	0.5-3.5	–	Marble bed	2	–	70	8/72	1
Paddys Run 6 Louisville Gas & Electric	Lime (Carbide)	Combustion Engineering	R	65	3.5-4	ESP	Marble bed	2	–	80	4/73	1
Philips Duquesne Light	Lime with MgO	Chemico	R	410	1-2.8	ESP	Three 1-stage venturi trains One 2-stage venturi train	4	30	83 90	7/73	1
EPA Shawnee Test Facility Tennessee Valley Authority	Lime and limestone	Chemico UOP	R	10 10	3-5	(b)	Venturi/spray tower TCA	1 1	Varies with each test		4/72	1
Colstrip 1 Montana Power	Alkaline fly ash with lime	ADL/CEA	N	360	0.8	–	Venturi/spray tower	3	15&18	60	10/75	1
Colstrip 2 Montana Power	Alkaline fly ash with lime	ADL/CEA	N	360	0.8	–	Venturi/spray tower	3	15 & 18	60	7/76	1
Reid Gardner 1 Nevada Power	Sodium carbonate	ADL/CEA	R	125	0.5-1	Mechanical	Venturi/sieve tray tower	1	–	85	4/74	1
Reid Gardner 2 Nevada Power	Sodium carbonate	ADL/CEA	R	125	0.5-1	Mechanical	Venturi/sieve tray tower	1	–	85	1/74	1
Reid Gardner 3 Nevada Power	Sodium carbonate	ADL/CEA	N	125	0.5-1	Mechanical	Venturi/sieve tray tower	1		85	7/76	1
Eddystone 1A Philadelphia Electric	Magnesium oxide	United Engineers–Peco	R	120	2.5	ESP/Venturi	Venturi rod absorber	3	–	90	9/75	1
D.H. Mitchell II Northern Indiana Public Service	Sodium sulfite (Wellman Lord)	Davy Powergas–Allied Chemical	R	115	3.2-3.5	ESP/Venturi	Absorption tower	1	–	>90	8/77	5

(a) Converted from limestone boiler injection to conventional scrubbing in 1977
(b) Tests have been made with and without ESP

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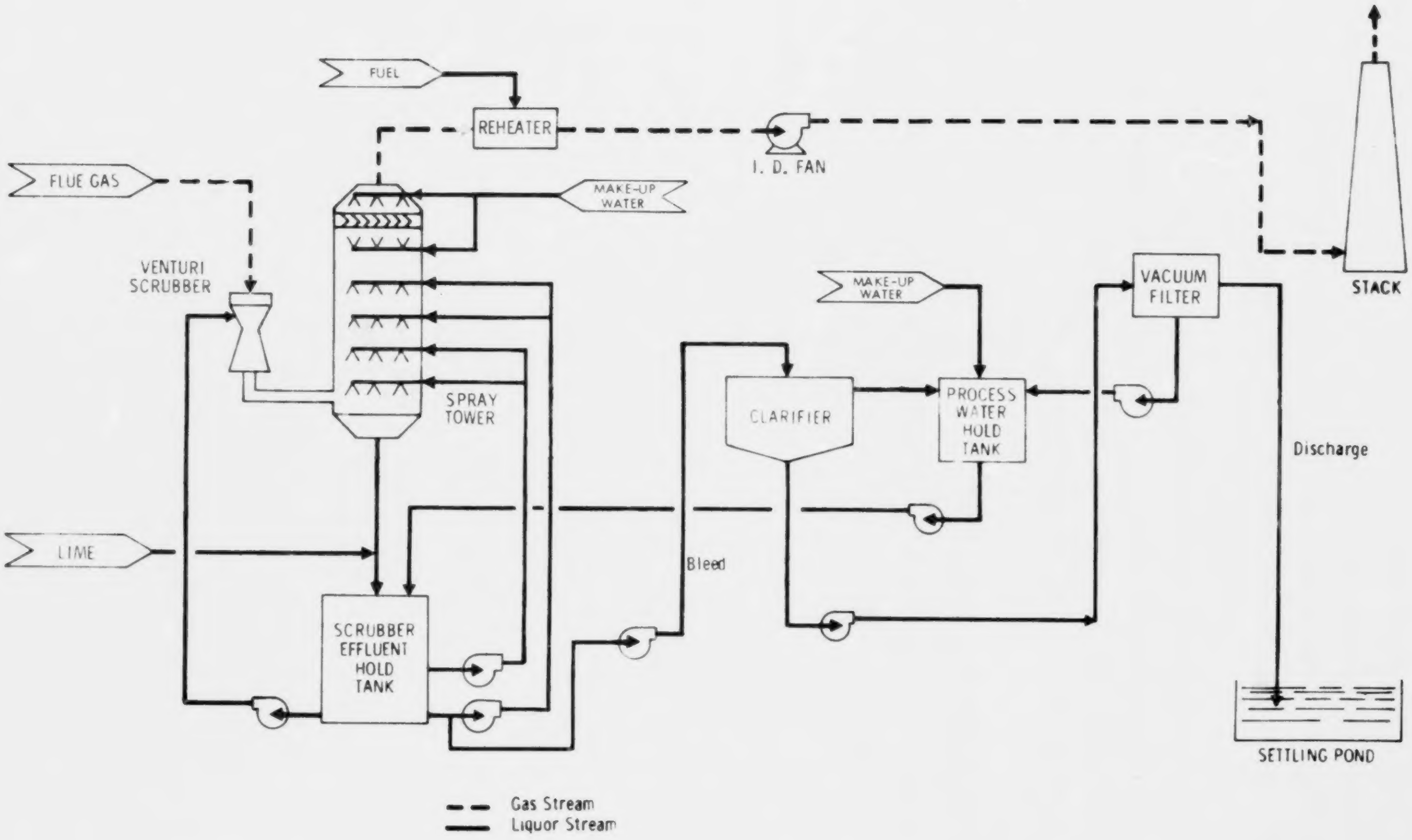
FUEL
REHEATER
I. D. FAN
STACK
FLUE GAS
VENTURI SCRUBBER
MAKE-UP WATER
SPRAY TOWER
MAKE-UP WATER
VACUUM FILTER
CLARIFIER
PROCESS WATER HOLD TANK
Discharge
LIME
Bleed
SCRUBBER EFFLUENT HOLD TANK
SETTLING POND
Gas Stream
Liquor Stream

Figure 3-1. Typical Process Flow Diagram for Lime/Limestone Scrubbing

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to a pond, thereby eliminating the filter. The clarifier may also be eliminated by sending the still greater volume of bleed stream directly to a pond. As the solids settle to the bottom of the pond, the clear liquor on top may be returned to the scrubber loop.

Makeup water is added to the system to replace the water evaporated into the flue gas and the water entrained in the waste stream. The water is added as mist eliminator wash, through pump seals, and through the lime slaker or feed makeup.

Process Status and Background

Operational lime and limestone FGD systems on U.S. coal-fired utility boilers are included in Table 3-1. In addition to the systems currently operating, there are 25 additional lime and limestone units under construction, totaling about 11,000 MW. Several full-scale operational systems have also been terminated. These include (Ref.1):

- Meremac 2, Union Electric — 140 MW limestone injection/wet scrubbing — experimental installation — abandoned
- Four Corners 5A, Arizona Public Service — 160 MW limestone scrubbing — test program completed and system shut down
- St. Clair 6, Detroit Edison — 133 MW limestone scrubbing — switched to low sulfur coal
- Valmont 5, Colorado Public Service — 50 MW limestone scrubbing — experimental — continuation of future FGD operations is contingent upon regulatory decisions
- Mohave 1A, Southern California Edison — 170 MW limestone scrubbing — test program completed and system shut down
- Mohave 2A, Southern California Edison — 160 MW lime scrubbing — test program completed and system shut down

Because they use naturally occurring materials and produce relatively insoluble wastes, and because their capital and operating costs are not as high as many other processes, lime and limestone scrubbing processes have received the most attention of all flue gas desulfurization methods. Significant improvements have been made over the past several years in the areas of reliability, variable load operation, system control, and sludge disposal techniques.

Process Design and Control Parameters

A number of system parameters must be monitored and controlled to ensure proper operation of lime and limestone FGD systems, to accommodate variations such as changing boiler load and inlet SO_2 conditions, to ensure proper mist eliminator operation, and to prevent scaling in the scrubber.

Changing Inlet SO_2 and Boiler Load. An FGD system design goal is to achieve a specified SO_2 removal efficiency for the most severe operating conditions anticipated, such as maximum boiler load (equivalent to maximum gas volume to be treated) and maximum sulfur content of the coal to be burned (equivalent to maximum SO_2 concentration in the gas). The removal efficiency is largely fixed by the hardware, as described in Section 2, and by the chosen level of operating parameters (see Section 4). Once these parameters are set, the system should achieve a certain minimum SO_2 removal at maximum boiler load and SO_2 concentration and even operate at higher absorption efficiencies for less severe conditions.

Examples that show the response characteristics of 10 MW scrubbers under long-term testing at the EPA Alkali Scrubbing Test Facility are helpful in interpreting the effect of these process changes.

Figures 3-2, 3-3, and 3-4 are presented to illustrate operating characteristics with limestone using a TCA and lime with a venturi and spray tower in series.

Figure 3-2 (Reference 6) shows operation of a small TCA with limestone. The gas flow rate was constant at 2.6 m/sec, or 8.6 ft/sec (22,600 Nm^3/hr, or 20,500 acfm at 300°F) and the L/G was fixed at 9.8 l/m^3 (73 gal/mcf). Inlet SO_2 concentration varied from about 2,000 to 4,500 ppm with the normal variation of coal sulfur content. The SO_2 removal efficiency was held between 75 and 88 percent by varying the limestone feed rate to compensate for the varying inlet SO_2. Removal efficiency tended to fall off during periods of high inlet SO_2. This effect is described in Section 4.

Figure 3-3 (Reference 7) shows lime scrubbing operation for a small venturi and spray tower in series. During this test the gas flow rate through the scrubbers varied with the changing boiler load while the liquor flow rate remained constant. During operation the gas rate varied from 18,700 to 38,600 Nm^3/hr, or 17,000 to 35,000 acfm (1.4 to 2.9 m/sec, or 4.5 to 9.4 ft/sec in the spray tower), while the liquor rates were 2,270 l/min (600 gpm) to the venturi and 6,060 l/min (1,600 gpm) to the spray tower. The effect of this type of operation is that when the gas flow is reduced the L/G increases (up to 16 l/m^3, or 117 gal/mcf for the spray tower) and absorption efficiency is enhanced for these types of scrubbers (reducing the gas velocity in other types of scrubbers may cause the efficiency to deteriorate – see Section 4). During operation the inlet SO_2 concentration ranged from 1,000 to 3,500 ppm. Therefore, the most severe operating conditions the FGD system had to deal with occurred when gas rate (boiler load) was at a maximum and inlet SO_2 was at 3,500 ppm. In this situation the SO_2 absorption efficiency was about 70 percent. During less severe conditions higher absorption efficiencies were obtained, and ranged up to 98 percent at low boiler load and low inlet SO_2 concentration. This example of operation with changing gas flow rate and constant liquor rate is typical for

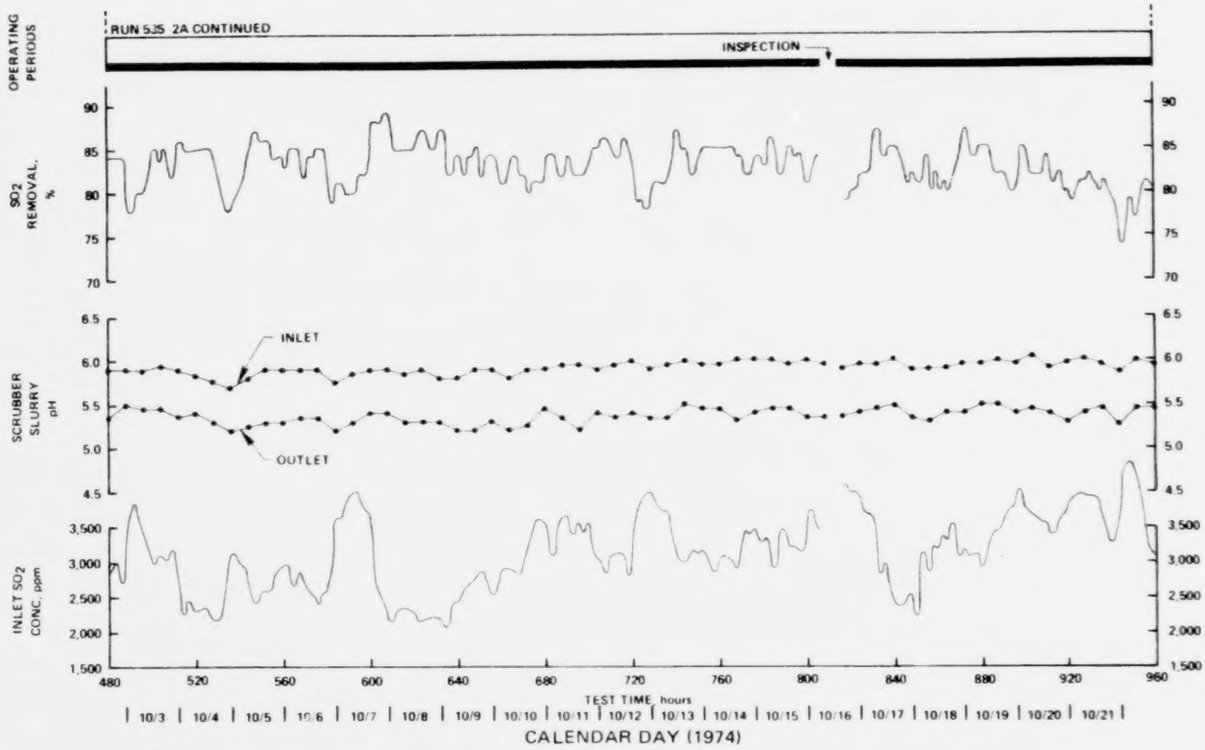

Figure 3-2. Operating Data from Shawnee Test Facility — TCA with Limestone

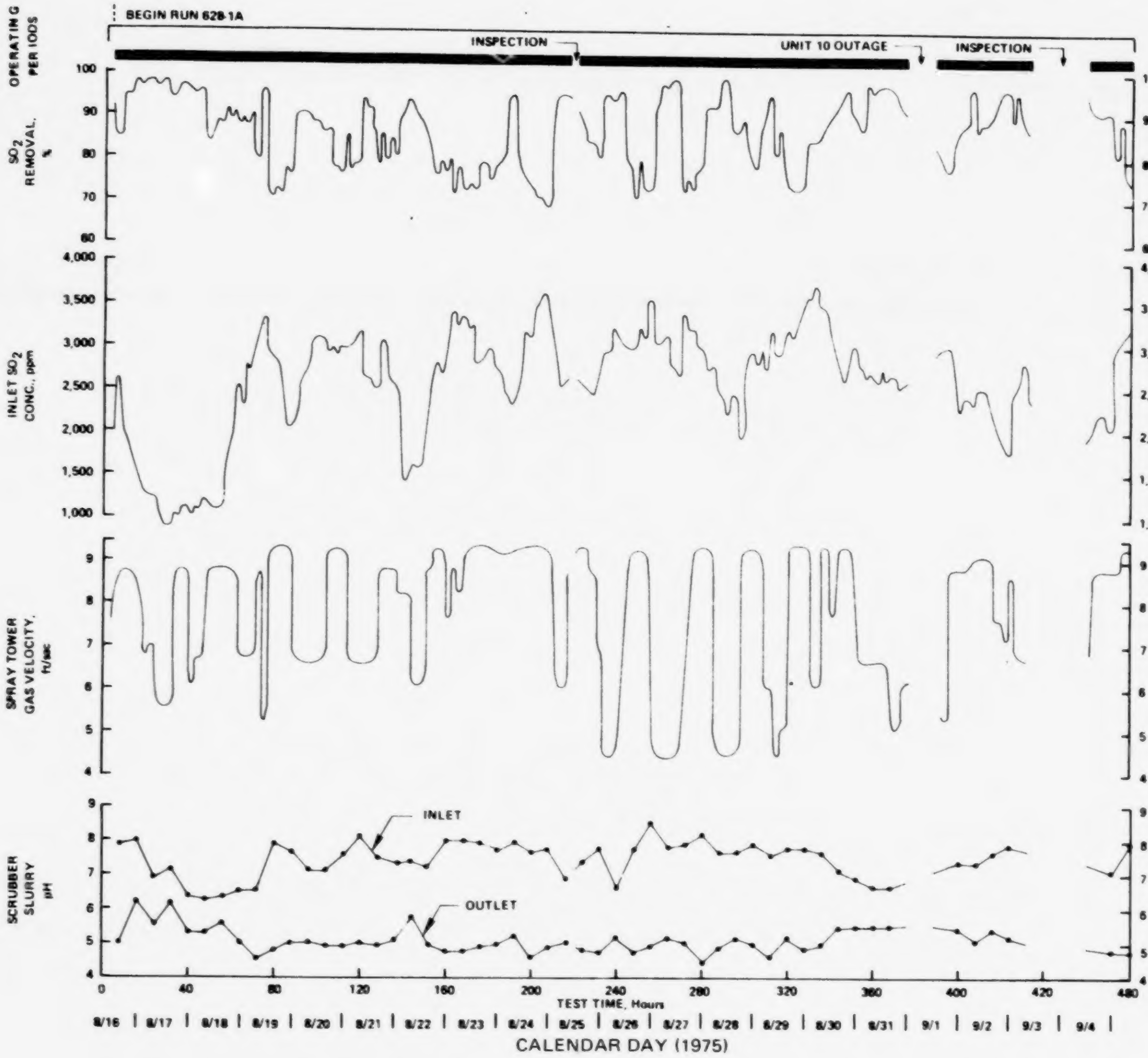

Figure 3-3. Operating Data from Shawnee Test Facility — Venturi/Spray Tower with Lime

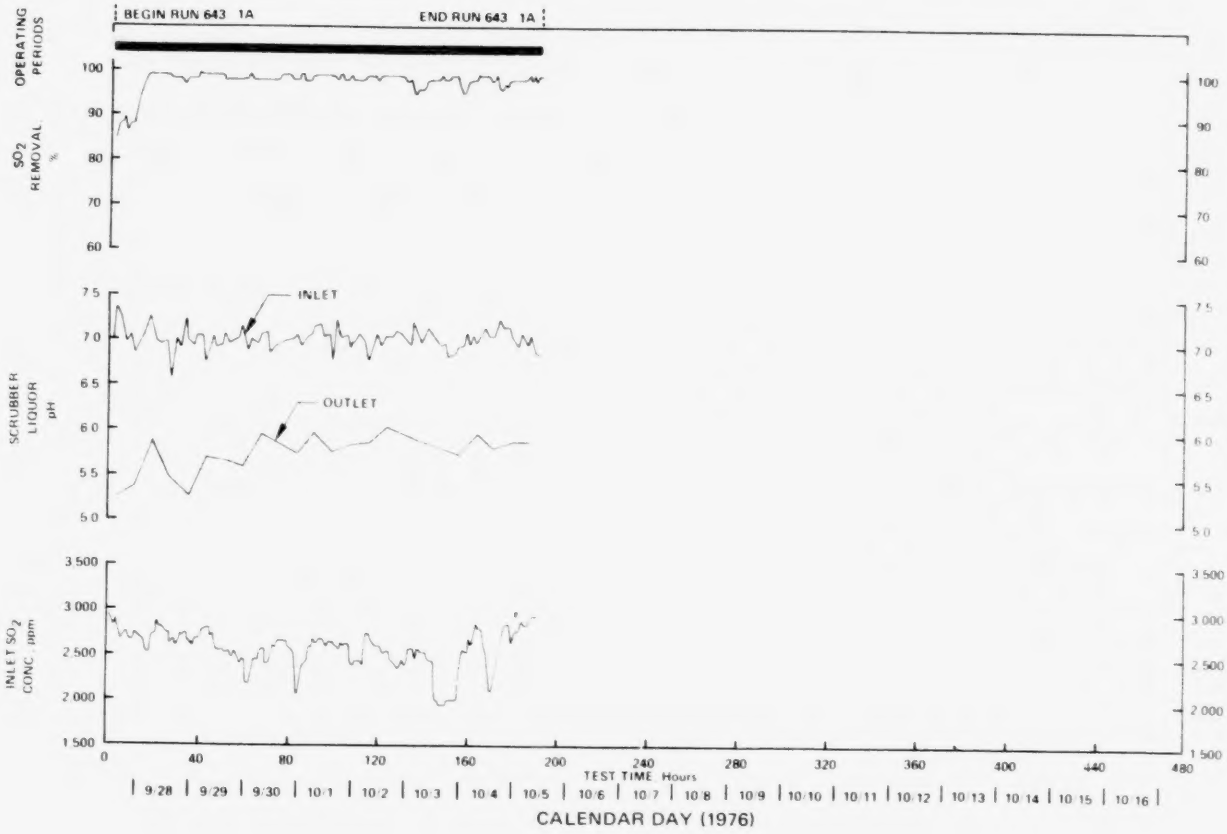

Figure 3-4. Operating Data from Shawnee Test Facility – Venturi/Spray Tower with Lime and Magnesium

installations using constant speed pumps. During operation the alkali feed rate is continuously adjusted to compensate for the varying amounts of SO_2 absorbed.

Figure 3-4 (Reference 8) again shows lime scrubbing operation for the venturi and spray tower in series. The previous two figures have shown that SO_2 absorption efficiency may vary over a wide range during normal operation. In Figure 3-4 the gas rate is 38,600 Nm^3/hr, or 35,000 acfm (2.9 m/sec, or 9.4 ft/sec in the spray tower) and the L/G's are 2.8 l/m^3 (21 gal/mcf) for the venturi and 15 l/m^3 (50 gal/mcf) for the spray tower. Magnesium was added to enhance the SO_2 removal efficiency (see Section 4). The inlet SO_2 ranged from 2,000 to 3,000 ppm while absorption efficiency ranged from 98 to 95 percent.

Control of Mist Eliminator Fouling. A mist eliminator is installed in the scrubber or downstream of the scrubber to catch mist carried over in the cleaned gas. This is done primarily to:

- Limit the emission of particulates to the atmosphere
- Prevent corrosion of downstream ductwork, fans, and stack lining
- Prevent plugging of inline reheater tubes and solids deposition on fan blades
- Reduce reheater energy requirements by removing water mist from the flue gas stream

The chevron mist eliminator is a commonly used style for FGD scrubbers. This device consists of a set of baffles that force the gas to flow in a zigzag pattern. As the gas passes through the mist eliminator, entrained mist is impacted on the blades and coalesces to form larger droplets which drain into the scrubber or into a collection trough. Other concepts have been developed including radial-vane, wet precipitation sections, two stages of chevrons, and many proprietary variations of the basic chevron design. In FGD scrubbers the mist eliminators have experienced slurry solids and chemical scale buildup in the narrow passages with resulting restriction of gas flow. Radial vanes may impart an undesirable swirling motion to the gas because of the shear effect. Many techniques have been employed to improve mist collection and minimize operational problems. These are summarized in Reference 3 and include:

- Washing. The mist eliminator can be washed with a spray of process makeup water or a mixture of makeup water and clarified liquor. It may be washed from one or both sides, and washing may be continuous or intermittent.

- Gas Velocity Reduction. The gas velocity may be reduced to 2 to 3 m/sec (7 to 10 ft/sec) by expanding the width of the mist eliminator section of the scrubber. This reduces the amount of entrained mist that reaches the mist eliminator.

- Wash Tray. Wash trays have been installed ahead of the mist eliminator to reduce the entrained slurry solids. The gas passes through clear liquor that is circulated across the tray, while entrained slurry solids tend to be captured.

- Horizontal Gas Flow. If the gas flow is reoriented to a horizontal direction, the mist eliminator can have a vertical configuration. This may offer superior drainage and be adaptable to a separate washwater recycle loop.

- Freeboard Distance. The distance the entrained mist must travel after leaving the final absorption stage before entering the mist eliminator can affect performance. At greater distances less entrained mist tends to reach the mist eliminator. About 1.2 meters (4 feet) is a minimum, but up to 5 or 6 meters (about 15 or 20 feet) has been employed.

- Slanting. A compromise between horizontal and vertical mist eliminator configurations is to use a slanted one in a vertical gas flow. The object is to improve drainage and reduce the potential for plugging.

- Mist Eliminator Design Factors. Many variations of the basic chevron design are in existence. Factors include the number of vanes (passes), spacing between vanes, angle of vanes, and various proprietary design details.

- Improved Limestone Utilization. During tests at the Shawnee facility, a correlation was found between alkali utilization and the accumulation of mud-type solids. Above about 85 percent alkali utilization the mist eliminators were kept clean with very infrequent (once per 8-hour shift) washing with fresh makeup water. An entirely clean system was maintained over an extended period using only 25 percent of the available makeup water. This correlation was confirmed using three different methods for improving alkali utilization, including reduced scrubber liquor pH, operating with 3 hold tanks in series, and during forced oxidation tests using staged scrubbing. The correlation was also confirmed at TVA's Colbert pilot plant.

Control of Chemical Scaling. The formation of hard scale on scrubber surfaces, particularly in the narrow passages of mist eliminators and on any packing material used, has been one of the main problems experienced with lime/limestone scrubbing. Calcium sulfite ($CaSO_3.1/2H_2O$) is precipitated from the chemical reaction of absorbed SO_2 with the slurry alkali. Calcium sulfate (gypsum, $CaSO_4.2H_2O$) is also precipitated when dissolved sulfite is oxidated by oxygen absorbed from the flue gas. With lime, calcium carbonate ($CaCO_3$) may be precipitated if the pH is too high when carbon dioxide (CO_2) in the flue gas reacts with the lime. Normally these reaction products crystallize on the suspended solids in the recirculating slurry. However, prevention of crystallization, particularly of calcium sulfate and calcium sulfite, on surfaces within the scrubber is a recognized problem in the design of lime and limestone processes.

Normally, as the scrubber liquor becomes supersaturated, crystallization will take place on the seed crystals present in the slurry. If the degree of supersaturation is allowed to increase above about 1.3 to 1.4, the precipitation will also occur on scrubber surfaces.

The techniques employed to prevent scaling typically include:

- Control of pH. If a limestone system is operated at pH's above 5.8 to 6.0, or if a lime system is operated above 8.0 to 9.0, there is a risk of sulfite scaling. The pH is controlled by adjusting the feed stoichiometry. Automatic control of feed by online pH sensors has been successful for lime systems but has generally proved impractical for limestone installations due to insensitive pH changes with limestone feed rate under the normal operating pH range. However, in the limestone system the feed can be automatically controlled by variations in flue gas flow rate and inlet SO_2 concentration.

- Hold Tank Residence Time. By providing retention time in the scrubber hold tank, the supersaturation of the liquor can be decreased before recycle to the scrubber. Typical residence times of 5 to 15 minutes have been demonstrated at Shawnee and used in some full-scale systems.

- Control of Suspended Solids Concentration. The degree of supersaturation can be minimized by keeping an adequate supply of seed crystals in the scrubber slurry. Typical levels in newer installations range from 5 to 15 percent suspended solids. Solids are generally controlled by regulating the slurry bleed rate, with occasional reset. Magnetic flow meters are effective monitoring devices, and slurry densities can be measured by various devices.

- Liquid-to-Gas Ratio. High liquid-to-gas ratios reduce scaling problems because the absorber outlet slurry is more dilute with respect to absorbed SO_2. Liquid rate is normally set for a constant flow, the L/G varying inversely with gas load.

SODIUM CARBONATE SCRUBBING

The absorption of SO_2 from flue gas may be accomplished by absorption in a clear water solution of sodium carbonate and sulfite.

Process Description

The process is shown in Figure 3-5. A solution of soda ash (Na_2CO_3) reacts with SO_2 in the absorber to form sodium sulfite/bisulfite. Both the reactant and the reaction products are completely soluble. If fly ash is removed prior to SO_2 removal, the absorber can be either a packed tower or a tray tower, high-efficiency devices with low pressure drops. If fly ash is not removed a venturi scrubber may be used for both particulate and SO_2 removal, with somewhat lower SO_2 absorption efficiency and

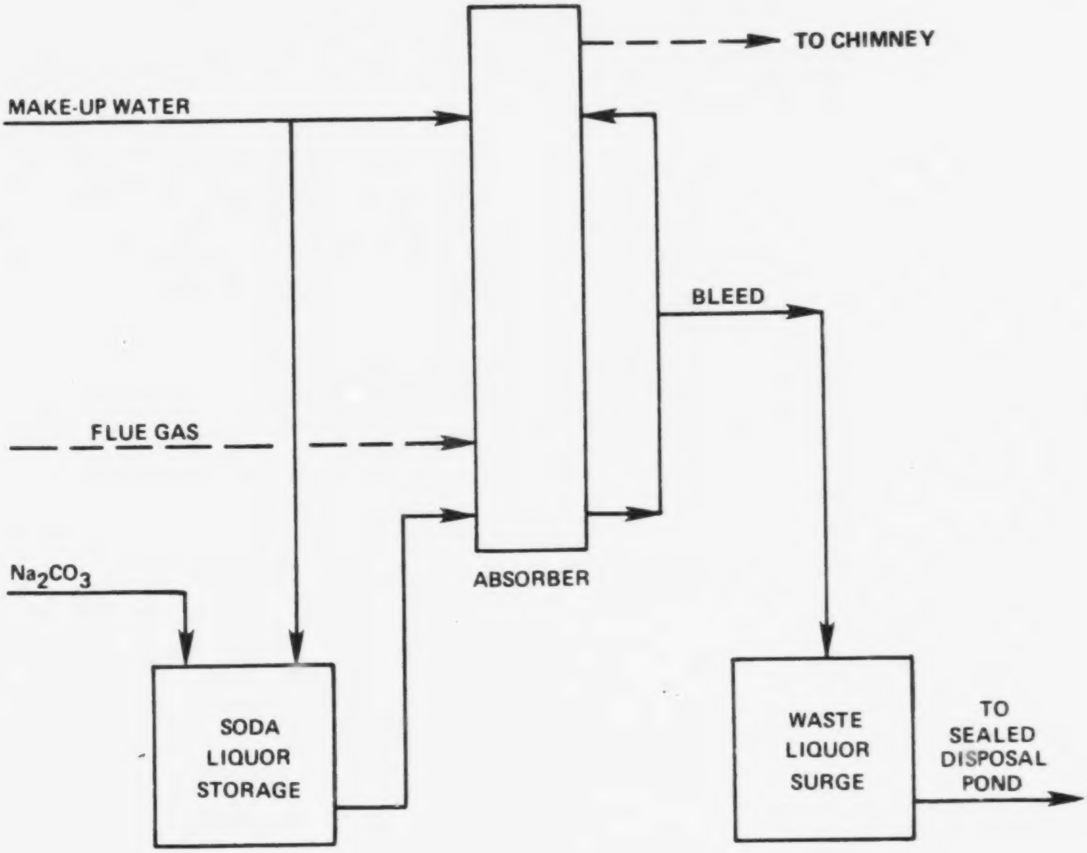
TO CHIMNEY
MAKE-UP WATER
BLEED
FLUE GAS
Na_2CO_3
ABSORBER
SODA LIQUOR STORAGE
WASTE LIQUOR SURGE
TO SEALED DISPOSAL POND

Figure 3-5. Simplified Process Diagram for Sodium Carbonate Scrubbing System

higher pressure drop. In either type of absorber, a recirculating liquid stream as well as fresh soda makeup is required to improve gas-liquid contact, which improves absorption efficiency. Because the sodium alkali is very reactive, the design L/G ratios can be low (in the 1.3 to 3.4 l/m^3, or 10 to 25 gal/Mcf range). The system responds rapidly to changes in SO_2 loadings, and the soda feed rate can be varied to match a pH signal from the absorber effluent, the pH changing with the SO_2 loadings.

Process Status and Background

A prototype unit, serving two industrial coal-fired boilers (equivalent to 25 MW) at the General Motors assembly plant in St. Louis has been in operation since 1972. Three 125 MW units at Nevada Power Company's Reid Gardner Station are operating with low sulfur coal (see Table 3-1) and a 520 MW installation is under construction at the Jim Bridger Station in Wyoming.

Process Design and Control Parameters

A stream of spent alkali solution, containing sodium sulfite/bisulfite equivalent to the amount of SO_2 removed, must be purged from the system to maintain chemical balance. The purge rate can be controlled by the liquid density. This purge stream, usually slightly acidic, is neutralized with more soda alkali before disposal. Process water makeup is required to compensate for the water evaporated in the flue gas and lost in the purge stream. Poor quality of this water (high calcium content) can lead to calcium scaling.

The spent alkali purge stream is normally discharged to a sealed (lined) evaporation pond for drying. Other proposed disposal methods include precipitation of the sodium salts and the recovery of sodium sulfate (salt cake) for sale.

The process consumes a premium chemical, either caustic soda or soda ash; therefore, its application is limited to small industrial boilers or utility boilers located near a source of the alkali. System advantages are simplicity, high absorption efficiency, and minimum capital cost. However, disposal of the spent alkali requires careful consideration.

DOUBLE ALKALI (SODA LIME) SCRUBBING

The double alkali process retains the high SO_2 absorption efficiency and nonscaling characteristics of the clear liquor sodium carbonate system. It avoids the sodium salt waste disposal problem by insolubilizing the scrubber effluent with lime precipitating calcium sulfite and sulfate. This likewise regenerates the costly sodium absorbent by displacing it with the less costly lime.

Process Description

The process is shown in Figure 3-6. As indicated, flue gas contacts clear soda solution in an absorber. The spent absorbent is mixed with lime slurry to precipitate the insoluble calcium salts and regenerate the absorbent. The calcium sludge is filtered, washed with water to recover the entrained soda solution, and discarded. Sufficient washwater is used to balance the water evaporated during the wet scrubbing of the hot flue gas.

Process Status and Background

Approximately 12 installations totaling 700 MW equivalent have been operated on oil-fired and coal-fired industrial boilers in the U.S. and Japan. Also, three installations totaling 1,050 MW on oil-fired utility boilers have operated in Japan, and 20 MW prototype testing has been carried out on a Gulf Power Company coal-fired utility boiler in Florida. Three full-scale units totaling 1,100 MW are scheduled for operation in the U.S. on coal-fired utility boilers (Refs. 1,9).

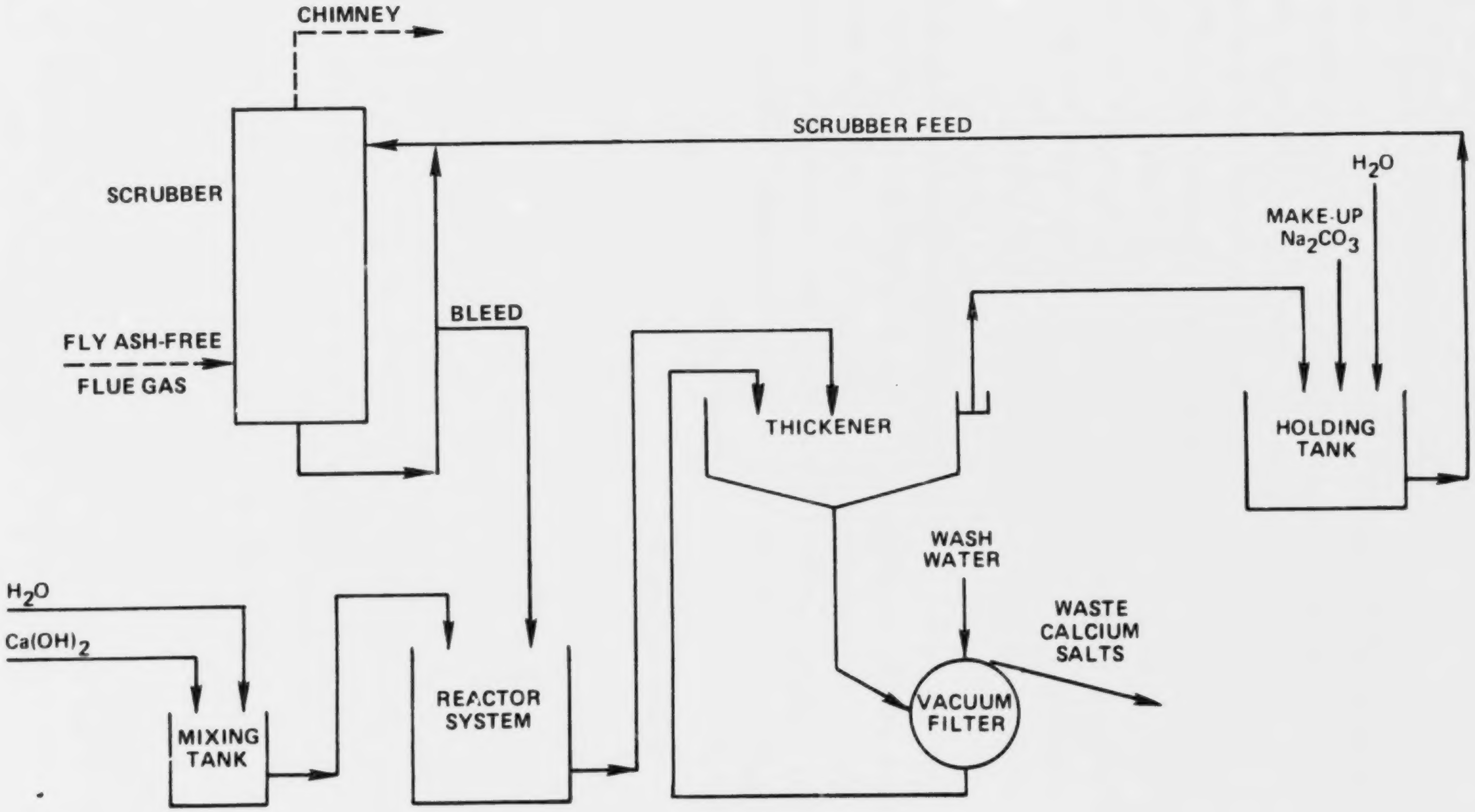
CHIMNEY
SCRUBBER FEED
SCRUBBER
H_2O
MAKE-UP Na_2CO_3
BLEED
FLY ASH-FREE FLUE GAS
THICKENER
HOLDING TANK
WASH WATER
WASTE CALCIUM SALTS
H_2O
$Ca(OH)_2$
REACTOR SYSTEM
VACUUM FILTER
MIXING TANK

Figure 3-6. Simplified Process Diagram for Double Alkali System

Process Design and Control Parameters

One of the main problems associated with the double alkali process is the difficulty in regenerating sodium sulfate because it does not react well with hydrated lime in the presence of sodium sulfite. Sulfate is formed by the oxidation of sulfite in the absorber. Oxidation can be minimized by using a concentrated absorbing solution. It is reported that as long as oxidation does not exceed 25 percent, the sulfate ion formed can be coprecipitated as calcium sulfate in the calcium sulfite crystals, thus releasing all the sodium for recycle.

Where oxidation cannot be controlled to below 25 percent (as with low-sulfur coal and/or high O_2/SO_2 ratios), a dilute absorbing solution can be used to overcome the detrimental effect of excessive oxidation. For dilute systems, in the absence of sulfite, the sulfate reacts well, regenerating NaOH for recycle to the absorber. However, the dilute mode has the following process disadvantages:

- A higher L/G ratio
- An extra softening step for the regenerated solution to prevent scaling (added cost)
- Higher lime requirement

The concentrated mode of operation, should oxidation exceed 25 percent, has the following process disadvantages:

- Soda makeup is increased to compensate for the purge of unregenerable sodium sulfate
- The sodium sulfate discarded in the waste sludge may result in brackish groundwater

In any one application, these factors must be weighed to determine the most suitable mode of operation. The concentrated mode is expected to be simpler and less costly and would be preferred for applications in which oxidation is expected to be relatively low. A concentrated mode would be indicated for high sulfur coals.

MAGNESIUM OXIDE SCRUBBING

The magnesium oxide scrubbing process is a wet slurry scrubbing recovery process. It differs from the similar calcium oxide (lime) process in that: (1) magnesium sulfite produced can be readily calcined to recover SO_2 in concentrated form, and (2) during calcining, the magnesium oxide is regenerated and can be recycled to the process. Thus, there is no sludge to dispose of, and only a small amount of chemical makeup is required. This is in contrast to the calcium-based scrubbing processes, which require a large quantity of chemical makeup, and disposal of large volumes of sludge. Although it is a slurry scrubbing process, scaling of absorbers in this system has not been a problem because magnesium sulfite is 300 times more soluble than the corresponding calcium sulfite, and magnesium sulfate is 120 times more soluble than calcium sulfate at the normal scrubbing temperatures.

Process Description

A simplified flow diagram illustrating the basic process steps of the magnesium oxide system offered by Chemico-Basic is shown in Figure 3-7. For coal-fired boiler application, fly ash must be removed from the flue gas before SO_2 absorption to prevent buildup of inert solids. Absorption of SO_2 takes place in a suitable contactor (such as a venturi) in which the flue gases are scrubbed with a circulating slurry containing about 10 percent solids by weight. Magnesium oxide (MgO) reacts with SO_2 to form magnesium sulfite hydrates. Some of the sulfite is oxidized to sulfate.

The magnesium sulfite/sulfate crystals are withdrawn in a bleed stream from the absorber slurry loop to a centrifuge. The wet cake from the

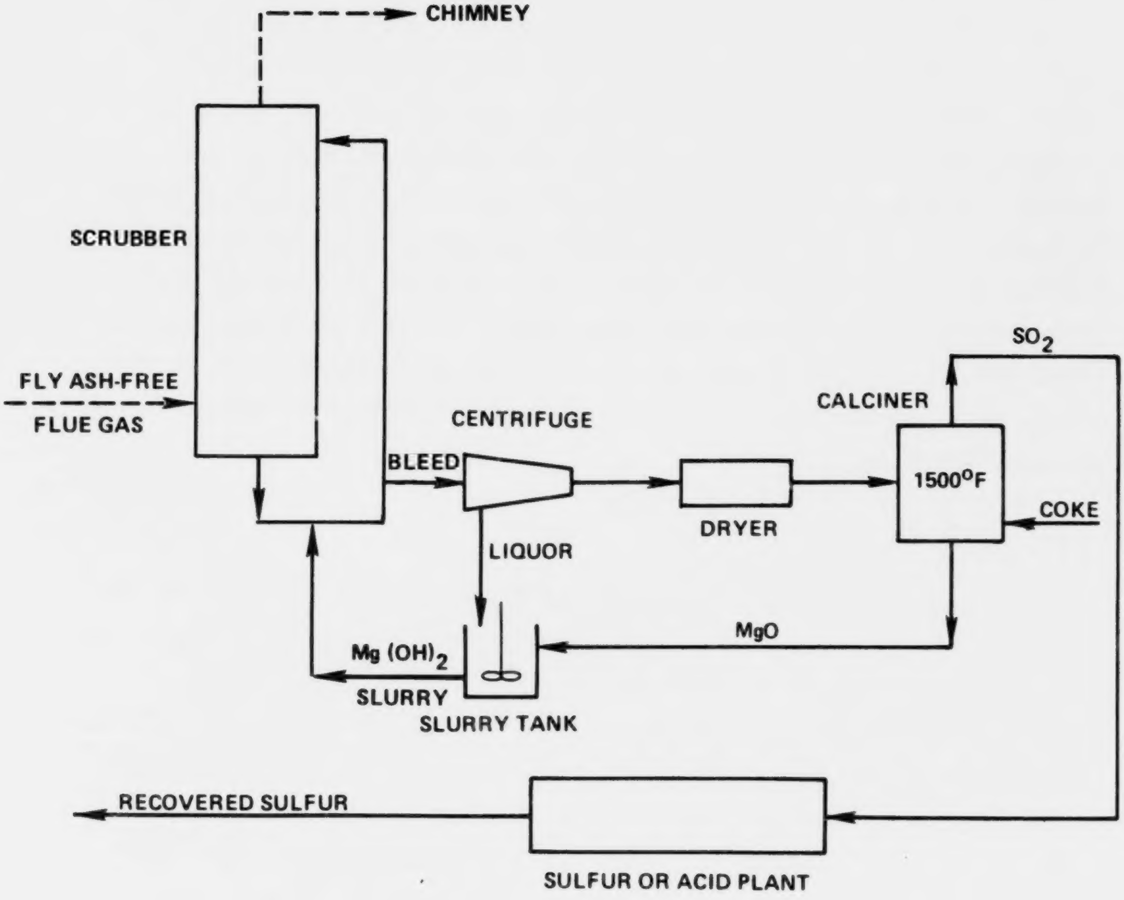
CHIMNEY
SCRUBBER
FLY ASH-FREE
FLUE GAS
CENTRIFUGE
BLEED
LIQUOR
DRYER
CALCINER
1500°F
SO2
COKE
MgO
Mg (OH)2
SLURRY
SLURRY TANK
RECOVERED SULFUR
SULFUR OR ACID PLANT

Figure 3-7. Simplified Process Diagram for Magnesium Oxide Recovery System

centrifuge is dried, and the remaining liquor is recycled to the absorber. The dried crystals are conveyed to a calciner where they are combined with a small amount of reducing agent (coke or carbon) and calcined under a reducing atmosphere at about 760° to $870^{\circ}C$ ($1,400$ to $1,600^{\circ}F$) to produce SO_2 and MgO. The reducing agent is required to reduce the sulfate only. The calciner exit gases contain between 10 and 16 percent SO_2 by volume on a dry basis.

After dust removal, the SO_2-rich stream may be used for sulfuric acid or sulfur production. The MgO, together with any required makeup, is recycled to the scrubber slurry makeup system (slaker).

Process Status and Background

In the U.S., two full-scale installations have been operated, one on an oil-fired boiler and the other on coal. A third full-scale unit is undergoing shakedown operation. These are summarized below.

A Chemico-Basic MgO scrubber at Boston Edison's Mystic Station went onstream in 1972 (Ref.10); the power plant generally operated on 2 percent sulfur fuel oil. The product — magnesium sulfite and sulfate — was shipped to a chemical plant for regeneration of the magnesia and production of by-product H_2SO_4. MgO was delivered to the plant in a very fine pulverized state (90 percent passing a 325-mesh screen). The SO_2 removal efficiency was higher than 90 percent. MgO losses averaged around 10 percent per cycle, although it was felt that with proper system control this could be reduced to about 5 percent while operating satisfactorily at steady-state conditions with regenerated MgO.

The major problem areas were: (1) formation of fine trihydrate sulfite crystals in the scrubber, which are more difficult to handle than the coarser hexahydrate crystals; (2) dust emission from the dryer; (3) erosion of pumps, valves, and piping; and (4) excessive wear on the internal parts of the centrifuge. The dryer dust problem was solved by rerouting the dryer off-gas to the scrubber inlet rather than the stack, but this prevented use of the dryer flue gas for stack gas reheat. The unit was operated during 1972-1974. Since then the utility has received a variance to burn higher sulfur oil.

Because Mystic Station was oil-fired, it did not provide information on the effects of fly ash on system operability. Fly ash must be efficiently removed ahead of the MgO scrubber to prevent buildup in the recycled material. An MgO scrubbing installation at Potomac Electric Power Company's Dickerson Station, started up in 1973, treating half the flue gas from a 190 MW coal-fired boiler. A two-stage venturi scrubber was used: the first stage removed the fly ash and the second stage removed the SO_2. Reheat was accomplished by mixing scrubbed flue gas with unscrubbed flue gas (Ref. 11).

The system operated intermittently from 1973 through 1975. Particulate removal was in excess of 99 percent; SO_2 removal efficiency ranged from 88 to 96 percent, depending upon gas flow rate. The MgO was operated through only one cycle. Although scrubber availability was not all that was desired due to problems with logistics in supplying raw materials (MgO), and to mechanical problems mainly attributed to underdesign in such areas as piping, slurry pumps, and other auxiliary equipment, it was felt that the basic concept and design have been proven to be feasible (Ref. 12).

The United Engineers system differs from the Chemico-Basic system in that: (1) the operating pH is lower (about 6) and the active absorbent is magnesium sulfite instead of magnesium oxide; and (2) magnesium oxide is added to the absorber effluent tank instead of the absorber (where it reacts with magnesium bisulfite to form magnesium sulfite).

The United Engineers MgO system at Philadelphia Electric Company's coal fired Eddystone Station (120 MW) was operated briefly in 1975, then shut down to relocate the MgO regeneration portion of the process nearby. The system started up again in 1977 and is undergoing shakedown (Ref 1).

The scrubber train consists of a venturi particulate scrubber followed by a venturi rod absorber. The system is reported to be designed to realize 90 percent SO_2 removal from 2.5 percent sulfur coal (Ref. 13).

Process Design Considerations

The dried magnesium sulfite is quite stable and easy to transport. For this reason the regeneration of the absorbent can be carried out at any convenient location quite distant from the power plant, thus permitting the use of a large central regeneration facility to serve several flue gas cleaning locations.

SODIUM SULFITE (WELLMAN LORD) SCRUBBING

The Wellman Lord system is a clear liquor sodium alkali scrubbing process with thermal regeneration and sulfuric acid or elemental sulfur recovery.

Process Description

A flow diagram illustrating the basic process steps is shown in Figure 3-8. In the absorber, sodium sulfite reacts with SO_2 to form sodium bisulfite in solution. In the regeneration evaporator, sodium bisulfite is thermally decomposed to sodium sulfite (solid) and SO_2 (gas). The SO_2 is stripped with steam and passes from the evaporator to a cooler where most of the steam is condensed. A concentrated SO_2 stream (about 90 percent SO_2 and 10 percent H_2O) is fed to a sulfuric acid plant or to a sulfur plant. The sodium sulfite crystals are separated in a clarifier and redissolved in water for recycle to the absorber.

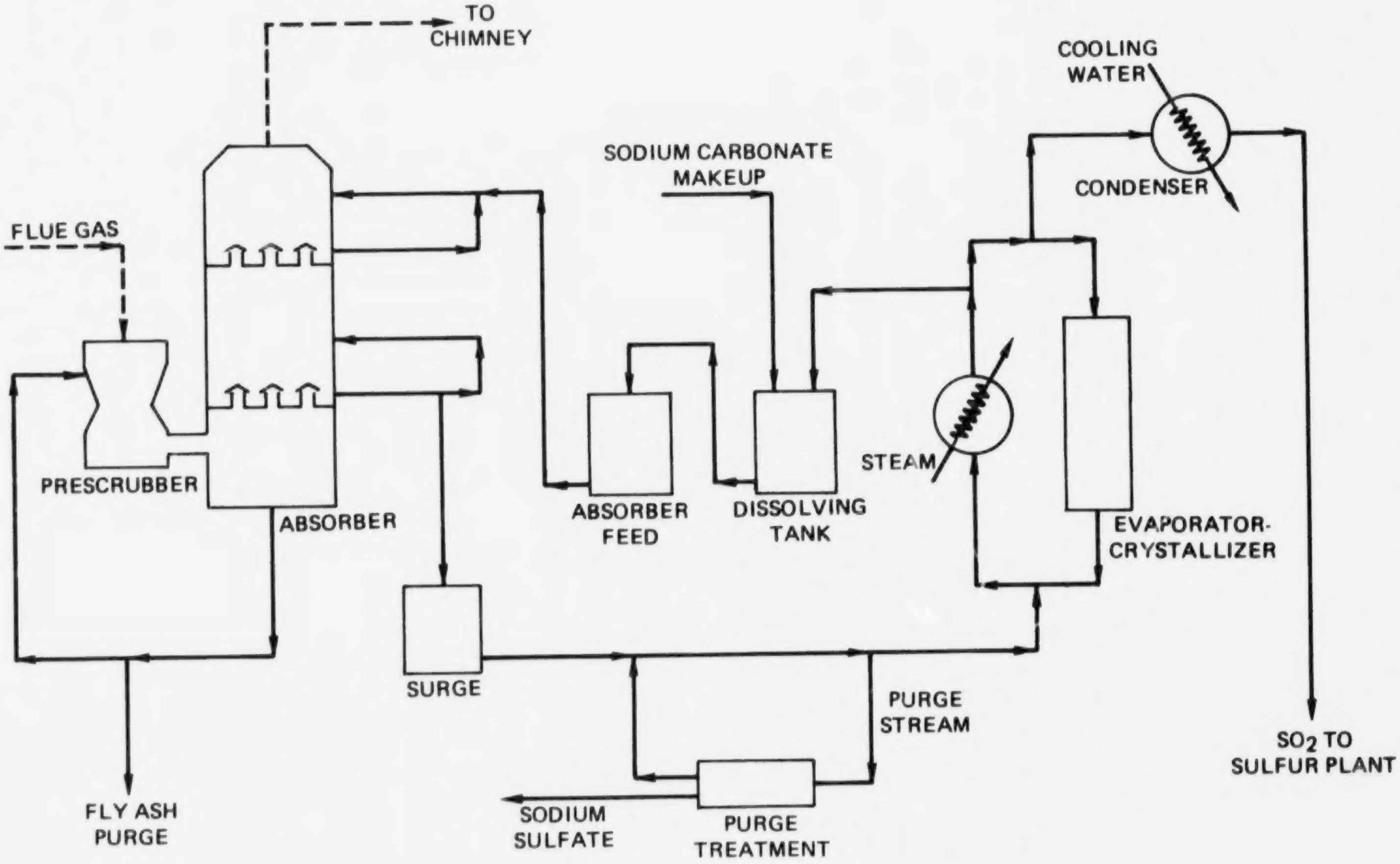
TO CHIMNEY
COOLING WATER
FLUE GAS
SODIUM CARBONATE MAKEUP
CONDENSER
PRESCRUBBER
ABSORBER
ABSORBER FEED
DISSOLVING TANK
STEAM
EVAPORATOR-CRYSTALLIZER
SURGE
PURGE STREAM
SO2 TO SULFUR PLANT
FLY ASH PURGE
SODIUM SULFATE
PURGE TREATMENT

Figure 3-6. Simplified Process Diagram for Wellman Lord Recovery System

Sodium sulfate is formed by oxidation of sodium sulfite in the system. This sulfate will not decompose in the regeneration evaporator, and must be purged from the system.

Process Status and Background

The process was developed originally by Wellman Lord, Inc., and is now licensed by Davy Powergas Company of Lakeland, Florida. A Wellman Lord system has been in operation on two oil-fired industrial boilers (35 MW each) in Japan since August 1971 with reported availability close to 100 percent. Two other full-scale installations on oil-fired boilers were started up in Japan during 1973: one is a 220 MW oil-fired utility boiler, and the other is a 125 MW equivalent industrial boiler. Both systems are reported to have been operating successfully.

Thus far the Wellman Lord installations in operation in the United States have been for treating tail gas for Claus plants and sulfuric acid plants. The first full-scale boiler installation in the United States is a retrofit at the coal-fired Mitchell station of Northern Indiana Public Service Company. This 115 MW unit is designed to recover elemental sulfur, using natural gas as the reducing agent. The unit has been completed and a comprehensive one year demonstration program is underway (Ref. 14). Two additional units totaling 715 MW are under construction for coal-fired utility boilers in New Mexico. The same power plant has signed a letter of intent for 2 additional units of 500 MW each.

Process Design Considerations

An important consideration in adapting the Wellman Lord process to coal-fired plants is the removal of particulare matter ahead of the absorber. Even with good particulate removal, the absorption liquor may have to be filtered to facilitate subsequent processing steps.

An additional concern is the oxidation of the sodium sulfite to sodium sulfate. From 5 to 10 percent or more of the total incoming sulfur is expected to be lost as soluble sodium sulfate in the purge stream. This means higher makeup costs, using soda ash or caustic soda. The purge stream must be disposed of in an environmentally acceptable manner. Methods are offered for recovering salt cake (Na_2SO_4) from this stream.

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Section 4

IMPLICATIONS OF REQUIRING 90 PERCENT OR GREATER SO_2 REMOVAL FOR NEW FGD INSTALLATIONS

This section discusses the changes in a demonstrated FGD system design of, for example, 80 percent SO_2 removal efficiency to obtain a capability of 90 percent or more. In summary, these design changes could involve:

- Increase in absorbent circulation rates
- Increase in scrubbing (contacting) stages or in contacting intensity (scrubber-type)
- More stringent requirements for uniform flow distribution
- Elimination of gas by-pass if used for reheat
- Increases in absorbent concentrations which may in turn require residence time or other changes to control enhanced scaling tendencies

If the increased SO_2 removal requirement is also accompanied by an increase in coal sulfur content, this change will intensify all of the above requirements. Other fuel related factors such as coal heating value, ash content and composition, must also be accommodated and with increasing difficulty as removal requirement rises.

LIME AND LIMESTONE SCRUBBING

This section presents data to illustrate the effects of process operating parameters on SO_2 removal efficiency for the lime and limestone scrubbing process. Most of the data are taken from test results at the EPA Alkali Scrubbing Test Facility located near Paducah, Kentucky (Ref. 1). The test facility is integrated into the flue gas ductwork of a 160-MW coal-fired boiler at the TVA Shawnee Power Station. Two

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parallel wet scrubber systems are in operation: a venturi/spray tower system and a turbulent contact absorber (TCA) system. Each system can treat approximately 10 MW (equivalent to 33,000 nm^3/hr, or 30,000 acfm at 300°F) of flue gas containing 1,500 to 4,500 ppm of SO_2.

The bulk of the data in this section is for limestone scrubbing on the TCA system. Additional data are presented in Appendix A for the TCA with the lime and the venturi and spray tower with lime and limestone operation. The Shawnee data are the result of short-term factorial tests 6 to 8 hours in length and long-term tests about one week in length. The short-term tests were aimed at determining SO_2 absorption efficiencies without regard necessarily for possible long-term scaling conditions that might be associated with the process conditions chosen. Therefore, these particular data are not to be interpreted as evidence of long-term reliable operation. However, during other portions of the Shawnee program adequate demonstrations have been made that these FGD systems can be operated for extended periods without scaling problems.

Approximate equations and nomographs for predicting SO_2 absorption efficiency for the Shawnee scrubbers are presented in Appendix B. For the TCA Shawnee data in this section, the SO_2 removal curves have been predicted by Equation B-4.

Effects of Process Variables on SO_2 Absorption Efficiency.

The effects of the following variables on SO_2 absorption efficiency are summarized in this section:

- Inlet SO_2 concentration
- Liquid-to-gas ratio
- Gas velocity

- Scrubber inlet pH
- Scrubber internals
- Magnesium addition
- Number of scrubbing stages
- Lime versus limestone

Inlet SO_2 Concentration. Figure 4-1 shows the effect of inlet SO_2 concentration on SO_2 removal for fixed design and operating conditions. Data have shown (Refs. 2, 3, 4) that in lime and limestone scrubbing over a wide range of inlet SO_2 concentrations, greater removal efficiency is realized for low inlet SO_2 concentrations than for high, for fixed design and operating conditions. This happens because the actual quantity of SO_2 removed per volume of gas treated goes down, reducing the loading on the absorbing liquor.

For high SO_2 concentrations, more SO_2 is actually absorbed by the liquor. For example, an absorber with 80 percent reduction from a gas containing 3,000 ppm SO_2 might absorb 90 percent from a 1,000 ppm gas, all things being equal. But in absolute terms, 2,400 ppm would actually have been removed from the higher SO_2 case, and only 900 ppm from the lower.

It was at one time thought that as inlet SO_2 concentration decreased, the driving force for absorption would decrease and high removal efficiencies would be more difficult. Clearly, as equilibrium SO_2 back pressure over the bulk liquid approaches the SO_2 gas concentration, this must pose a problem. In practice, however, the SO_2 back pressure over fresh lime and limestone scrubbing slurries is very low (less than 1 ppm), and scrubbers can be designed so that the low SO_2 concentration scrubbed gas exiting the scrubber is contacted by the freshest slurry (counter current or cross flows). For these reasons, removal efficiency greater than 95 percent has been achieved during tests from

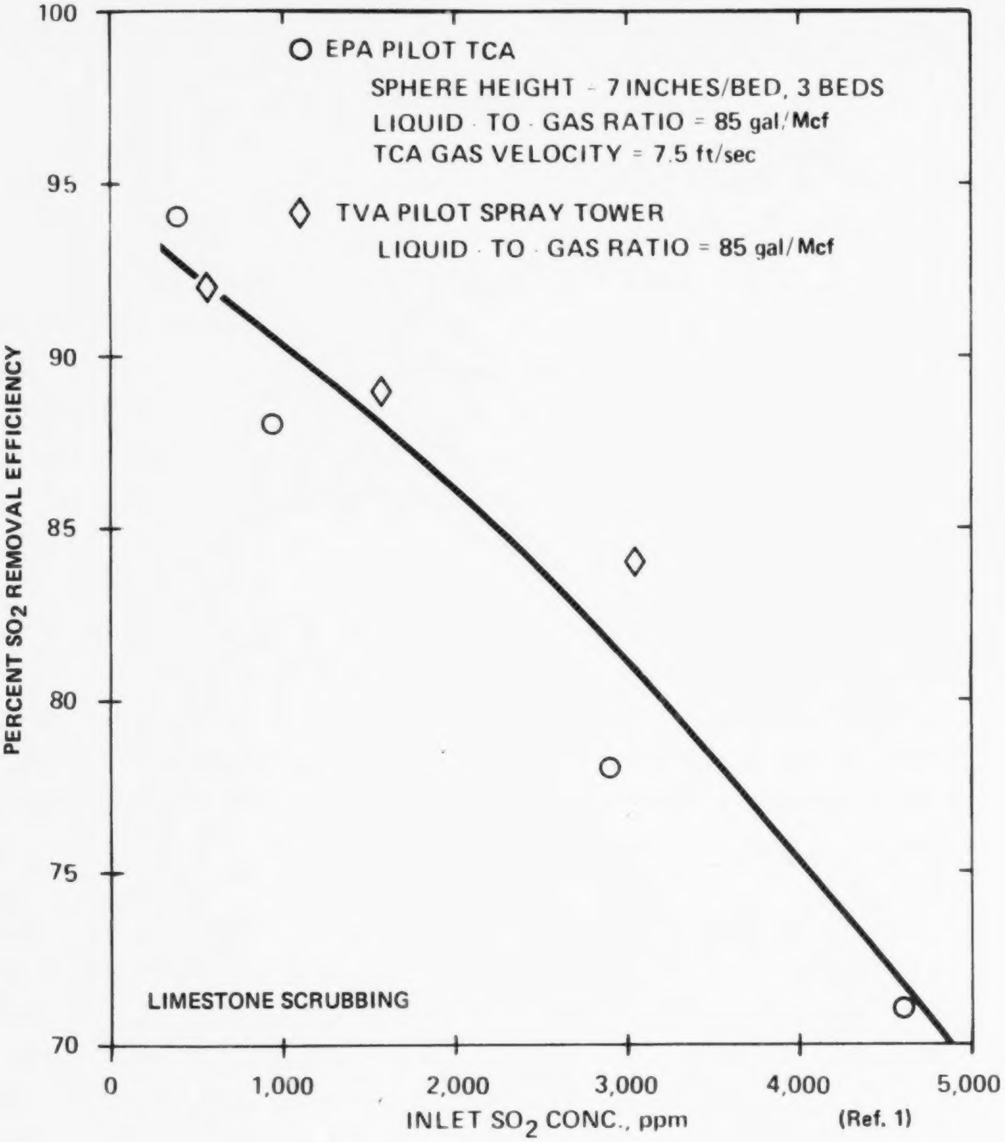


Figure 4-1. Effect of Inlet SO_2 Concentration on SO_2 Removal Efficiency for Fixed Design and Operating Conditions

flue gas that averaged only 200 ppm at the inlet, as shown in Figure 4-2 for a TCA system operated at the Southern California Edison Mohave Station (Ref. 5). The outlet SO_2 concentration in this case is less than 10 ppm. It is important to recognize that to obtain equivalent efficiency for 3,000 ppm SO_2 gas, an FGD system would need to absorb and process about 15 times as much SO_2 (2,850 vs 190 ppm). Achievement of comparable efficiencies at the high loading corresponding to 3,000 ppm SO_2 gas cannot be considered to have been demonstrated by the Mohave example. The gas SO_2 concentration, or coal sulfur content, must always be accounted for when extrapolating the performance of different installations.

Liquid-to-Gas Ratio. Figure 4-3 illustrates the effect of L/G on SO_2 removal efficiency in a TCA for limestone scrubbing. Similar effects are observed for lime and in the venturi and spray tower scrubbers as shown in Appendix A. Liquid-to-gas ratio is a major design variable in scrubbing systems. As the liquid pumping rate is increased, additional alkali is added to the scrubber and the gas-liquid interfacial area increases, resulting in greater absorption efficiency.

In most scrubber types, the benefits of increasing L/G for fixed scrubber condition can be designed into the system up to the point where problems such as flooding or nonuniform gas and liquor distributions occur. It can be seen from the way the curves flatten out at high L/G in Figure 4-3 that the incremental increase in pumping to improve the absorption from, say, 80 to 90 percent (an increase from about 50 to 80 L/G at a pH of 5.8) would be significantly greater than that required to improve it from 70 to 80 percent (an increase from about 35 to 50 L/G). Higher L/G could also be obtained by operating at lower gas velocity, at the expense of increased scrubber size.

An increase in the design L/G for an FGD system requires the use of larger pumps, pipes, and spray headers. Scrubber recirculation tanks

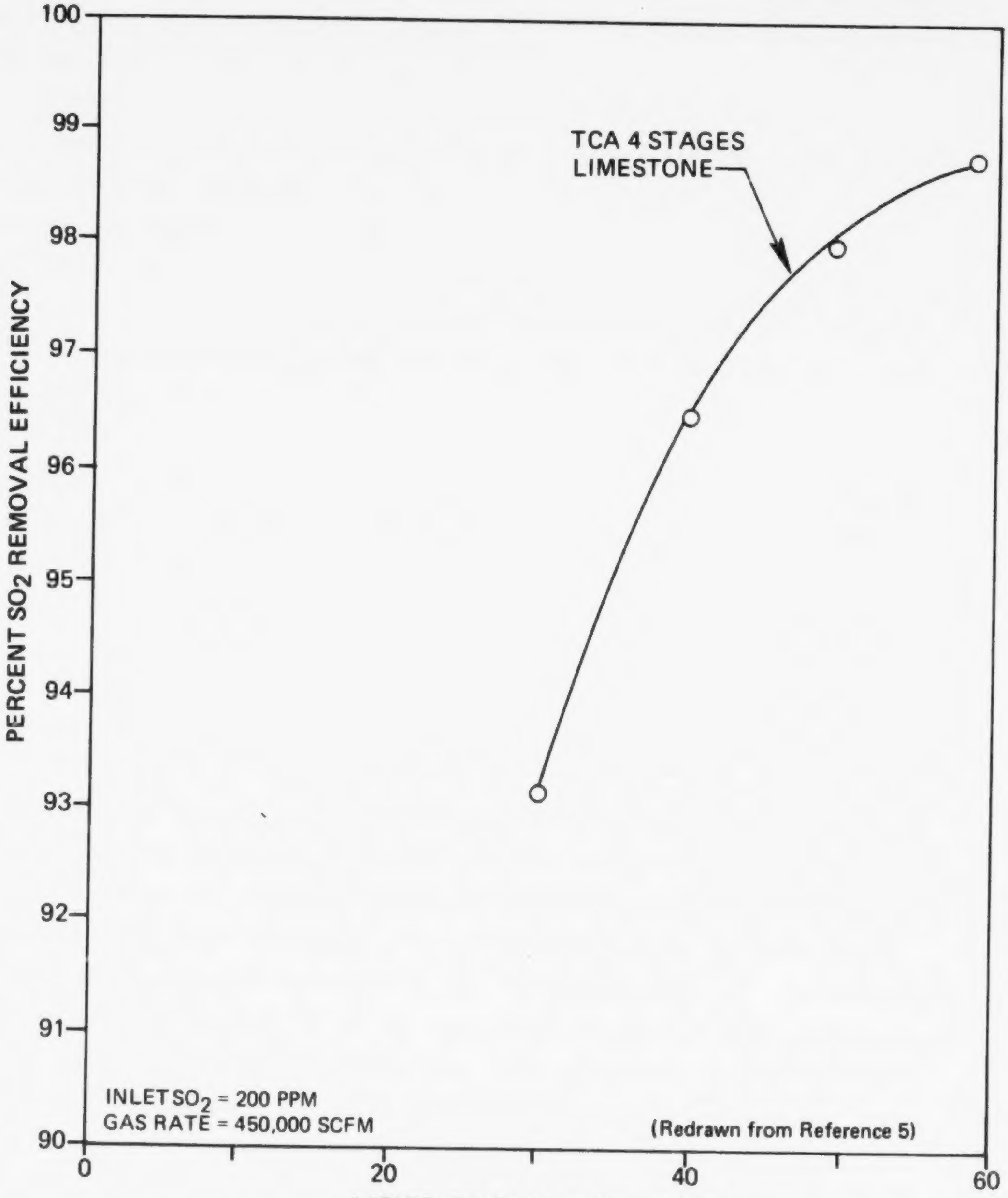
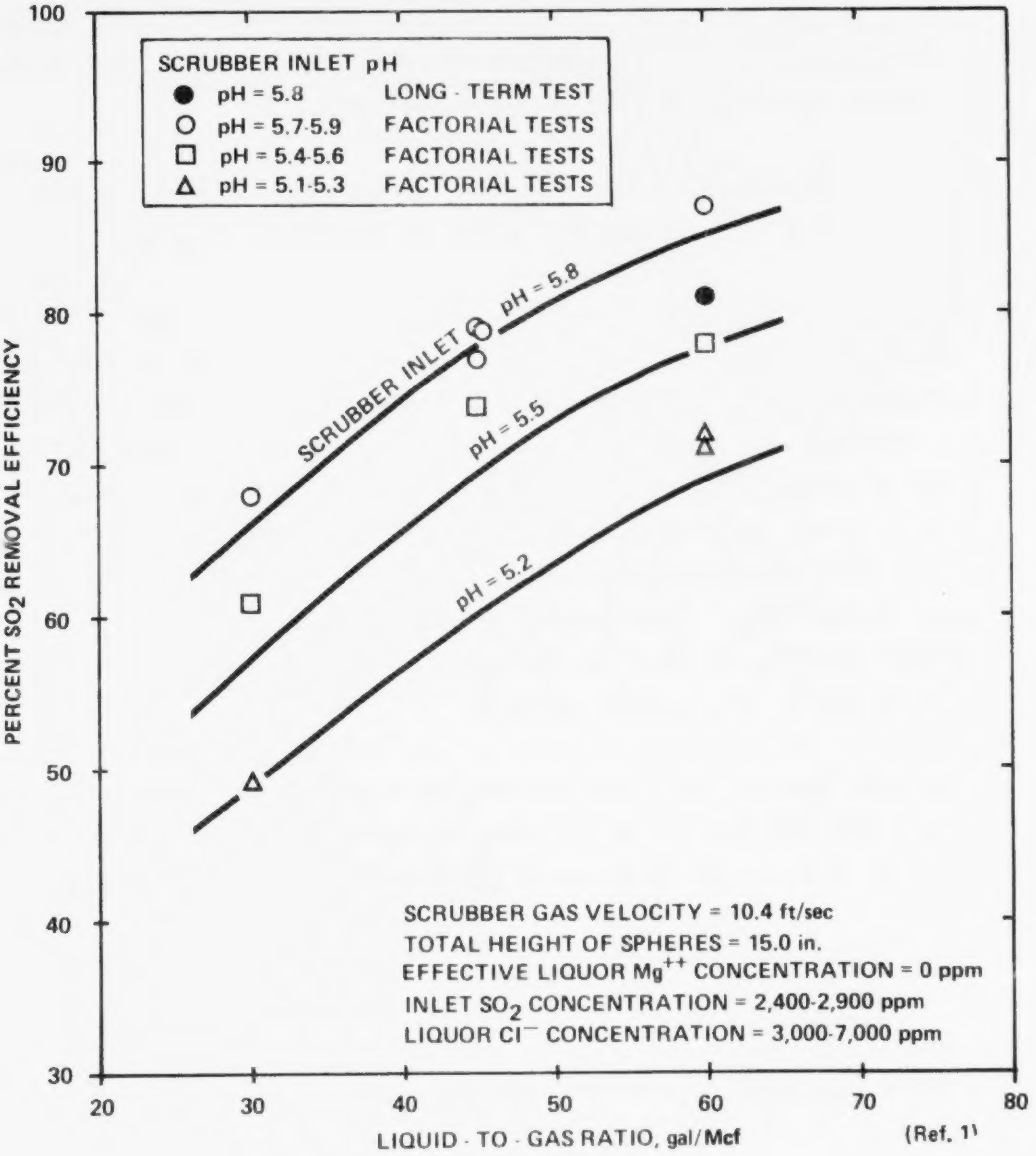


Figure 4-2. Effect of Liquid-to-Gas Ratio on SO_2 Removal Efficiency with Low Sulfur Coal at the Mohave Power Station



(Ref. 1)

Figure 4-3. Effect of Liquid-to-Gas Ratio on SO_2 Removal Efficiency — TCA with Limestone

must be larger to maintain design residence time for the increased L/G and to provide residence time for the increased SO_2 absorption. The energy penalty for increased L/G is shown in Figure 2-3. At higher L/G's, gas pressure drop may also increase in some types of scrubbers, causing additional penalty as shown in Figure 2-2.

Higher L/G's may tend to reduce scaling potential in the scrubber by reducing the SO_2 absorbed per gallon of absorbent.

Scrubber Gas Velocity. Figure 4-4 shows the effect of gas velocity on SO_2 removal efficiency in a TCA. The small velocity effect at a fixed slurry flow rate represents a trade-off between factors tending to increase efficiency with velocity (e.g., superior sphere agitation and increased liquid-gas contact) and the corresponding decrease in L/G. On the other hand, a scrubber such as the spray tower (Figure A-14) will have higher efficiencies at lower gas velocities, at a fixed slurry flow rate. In the spray tower, there are no internals and little liquid hold-up, so that the decrease in L/G with increasing gas velocity is not offset by other factors.

For cost reasons, scrubbers are designed to operate at maximum practical gas velocities, thereby keeping vessel diameters minimal. These maximum velocities are dictated by gas-liquid distribution characteristics of the scrubber and by the maximum allowable slurry entrainment that the mist eliminator can handle. Designers generally achieve the required L/G by increasing the slurry pumping rate rather than by decreasing the gas velocity. Designing for low gas velocities introduces larger and costlier scrubber systems (modules).

In specifying high removal efficiency, it should be kept in mind that some very efficient scrubbers (including certain tray towers) have poor modular turndown characteristics, so that absorption efficiency

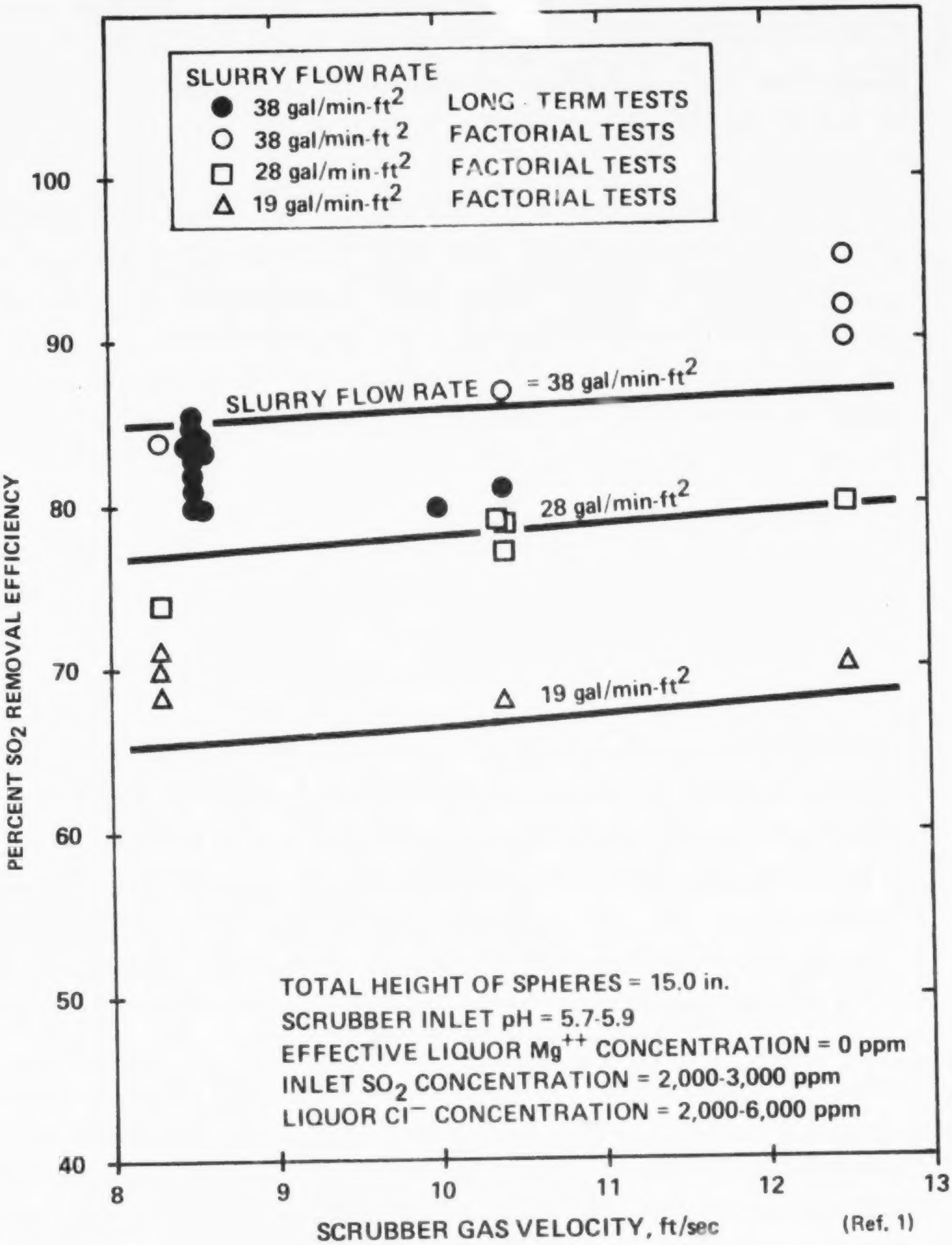


Figure 4-4. Effect of Gas Velocity on SO_2 Removal Efficiency — TCA with Limestone

falls off significantly as gas velocity decreases. Although such scrubbers may be designed to accommodate the normal boiler load turndown, they may lose efficiency below that range. One solution is to take modules off-line as load decreases.

Scrubber Inlet pH. Figure 4-5 illustrates the effect of scrubber inlet pH on absorption efficiency. Scrubber inlet pH is a measure of the lime or limestone feed stoichiometry (moles Ca added/mole SO_2 absorbed). As the alkali feed is increased (i.e., as pH goes up), the additional suspended alkali particles in the slurry lead to greater dissolution rates in the scrubber. Since lime and limestone dissolution tends to be limiting, increasing dissolution improves the efficiency.

If the lime or limestone processes are operated at too high a pH, calcium sulfite scaling (and calcium carbonate scaling in the lime system) occur in the scrubber. The actual maximum pH for reliable operation depends on such variables as dissolved salts (calcium chloride, for example) and liquid-to-gas ratio. For the systems explored at Shawnee, the maximum safe pH values are about 5.9 for limestone (corresponding to a stoichiometric ratio of about 1.4) and about 8.5 for lime (about 1.1 stoichiometric ratio). Slightly lower pH values of 5.8 and 8.0 are more comfortable and leave room for unintentional process excursions to higher values. It is not good practice to provide design latitude to achieve high efficiencies by operating at pH levels higher than these. Scale formation caused by operation at such levels may plug scrubber passages, preventing gas flows and increasing gas pressure drop. Dislodged scale may damage slurry recirculation pumps and plug spray nozzles.

As alkali feed stoichiometries are increased, the excess alkali reacts incompletely with absorbed SO_2 and is discarded, unused, with the waste sludge. For limestone scrubbing at a stoichiometric ratio of 1.4, only 71 percent of the limestone is used. Additional increase in stoichiometry

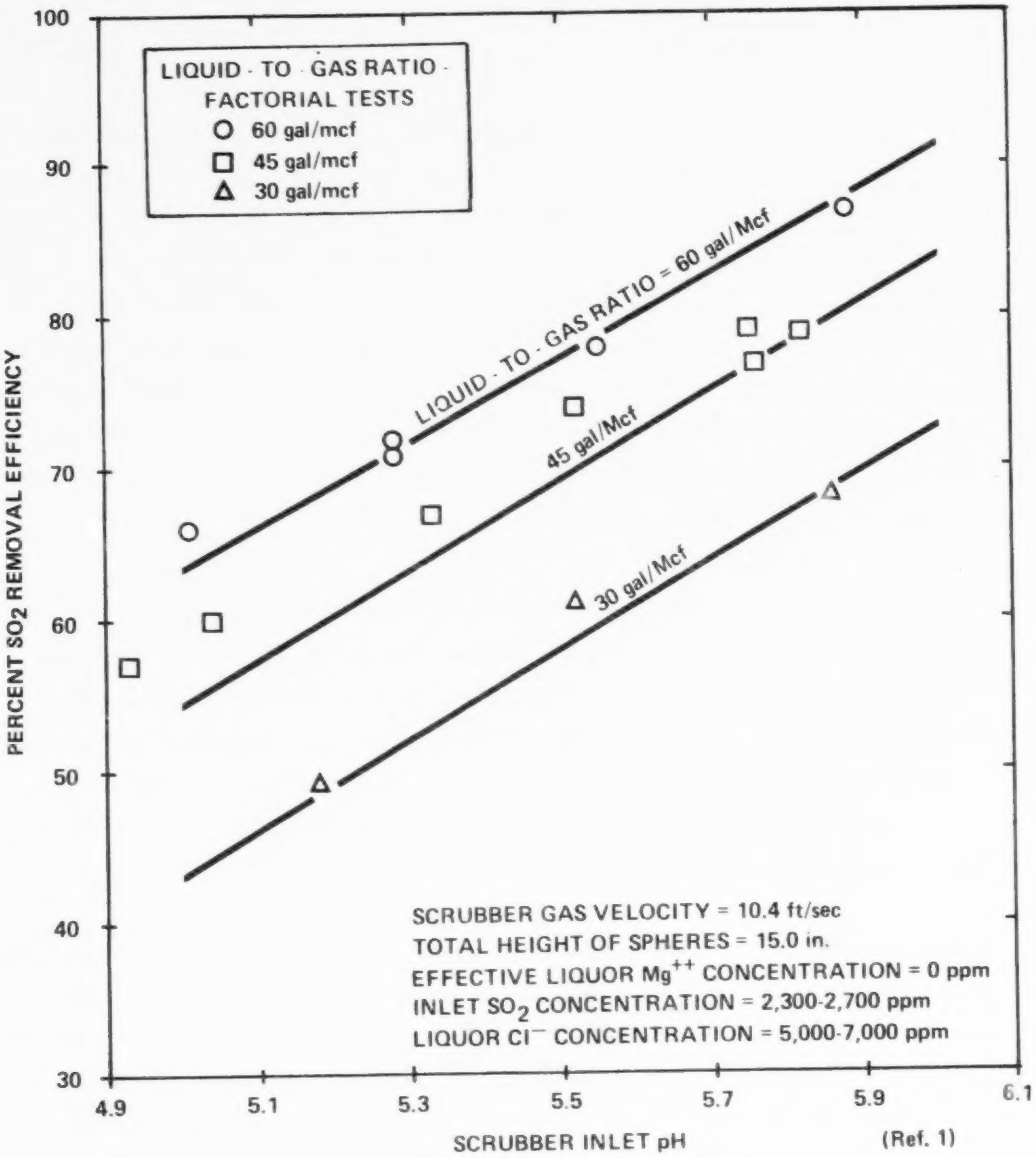


Figure 4-5. Effect of Scrubber Inlet pH on SO_2 Removal Efficiency — TCA with Limestone

wastes even more limestone and adds to the volume of waste sludge. This is another drawback to the use of increased pH (and stoichiometry to increase SO_2 removal.

Scrubber Internals. Figure 4-6 shows the effect of mobile packing height on SO_2 removal efficiency in a three-bed TCA. Increasing bed heights by adding more spheres increases slurry holdup in the scrubber. This has the combined beneficial effects of providing additional liquor holdup time for alkali dissolution and a greater gas-liquid contact area for improved efficiency. In the TCA, where sphere retaining grids are about four feet apart, the static sphere height per bed can be adjusted over a range of several inches. If greater removal is desired, the number of beds can be increased. Increasing the bed height, however, increases the gas pressure drop.

Other improvements in efficiency are achievable by increasing the packing height in fixed packing towers and by adding more trays in tray towers, although these types of scrubbers are not often used in slurry service because of their susceptibility to plugging. These improvements are made at the expense of increased gas pressure drop across the scrubber and, therefore, increased fan power requirements (see Figure 2-2).

Magnesium Addition. Adding soluble magnesium species to lime and limestone process slurries is beneficial to scrubbing efficiency. The alkaline magnesium compounds are highly soluble (in the form of magnesium sulfite and sulfate) compared to calcium-based compounds and build up to high concentrations in the scrubbing liquor. Since the absorbed SO_2 can react directly with the liquid-phase active magnesium species (magnesium sulfite), the absorption rate is not limited by the slower dissolution of lime or limestone suspended solids, and efficiency is improved. Depending on the magnesium concentration in the liquor, the absorption can be accommodated almost entirely by the magnesium, or only partially by the magnesium and the remainder by the lime or limestone.

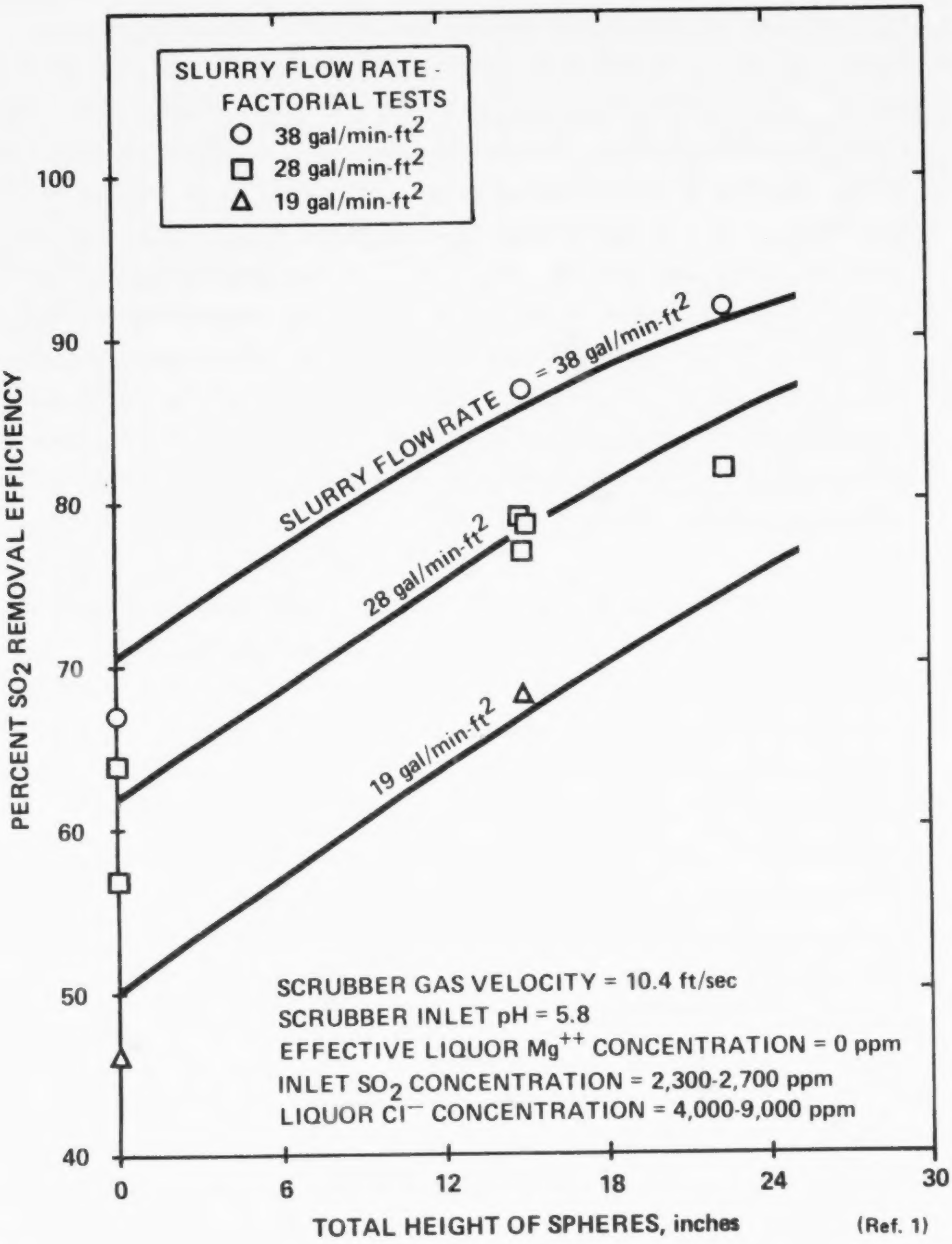
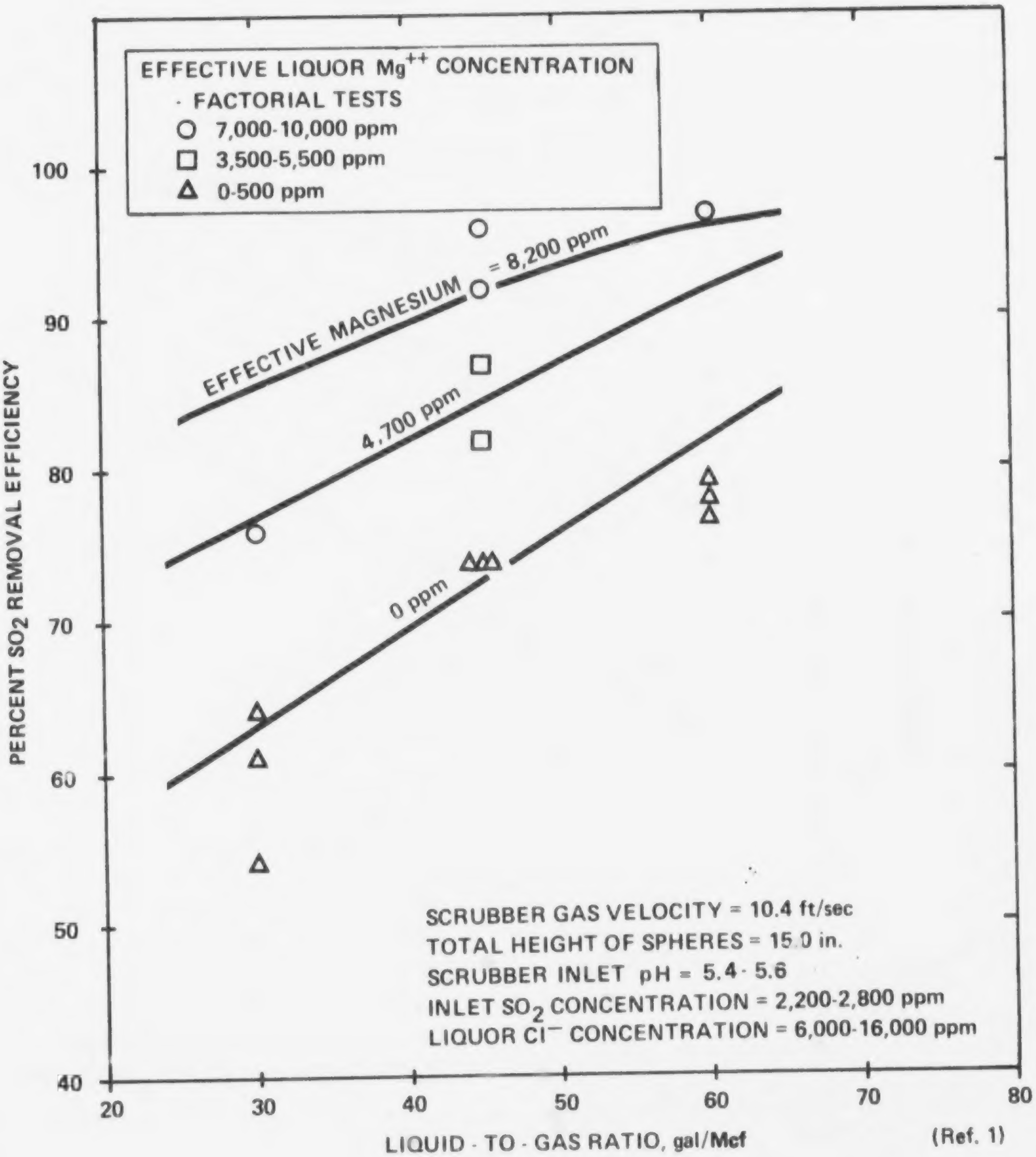


Figure 4-6. Effect of Bed Height on SO_2 Removal Efficiency — TCA with Limestone

Figures 4-7 through 4-9 show the combined effects of dissolved magnesium, L/G, and pH on SO_2 removal for the TCA with limestone. Magnesium can be added to the process as magnesium oxide or magnesium sulfate or, in a lime system, as dolomitic lime. In any case, the active portion of the dissolved magnesium enters the scrubber as magnesium sulfite, $MgSO_3$, and leaves the scrubber after absorption as magnesium bisulfite, $Mg(HSO_3)_2$. In the slurry hold tank the lime or limestone continues to dissolve and react with $Mg(HSO_3)_2$ to form the normal lime and limestone calcium-based reaction products (waste sludge) and to regenerate the active $MgSO_3$ for recycle to the scrubber. Soluble magnesium compounds are lost in the water portion of the sludge, requiring continued addition of makeup magnesium to the process. Also, the waste sludge must be treated or impounded to prevent groundwater contamination by the magnesium sulfate (Epsom salt).

The small amounts of chlorine in coal are converted to gaseous chloride in the boiler. The chloride is absorbed from the flue gas by wet scrubbing processes, and the resulting dissolved chloride ion can build up to appreciable concentrations in the scrubber liquor (2,000 to 16,000 ppm at the Shawnee facility, where chlorine concentration in the coal ranges from 0.03 to 0.3 percent and system water loop closure corresponds to a discharge stream containing 35 to 65 percent solids). The dissolved chloride neutralizes magnesium to an inactive form ($MgCl_2$) relative to the absorption process. To enhance absorption, magnesium concentrations must be increased above that amount made inactive by the chloride. This additional amount is referred to in Figures 4-7 through 4-9 as the effective liquor magnesium concentration.

The magnesium addition rate required to maintain a specified effective concentration is, therefore, a function of the FGD process water loop closure (amount of water lost in waste sludge) and the chlorine content of the boiler feed coal. At Shawnee, in the TCA operating with limestone, about 2 to 6.5 pounds of magnesium oxide per 100 pounds of limestone are required to maintain a 9,000 ppm effective magnesium concentration in the scrubber liquor.



(Ref. 1)

Figure 4-7. Effect of Liquid-to-Gas Ratio on SO_2 Removal Efficiency — TCA with Limestone and Magnesium

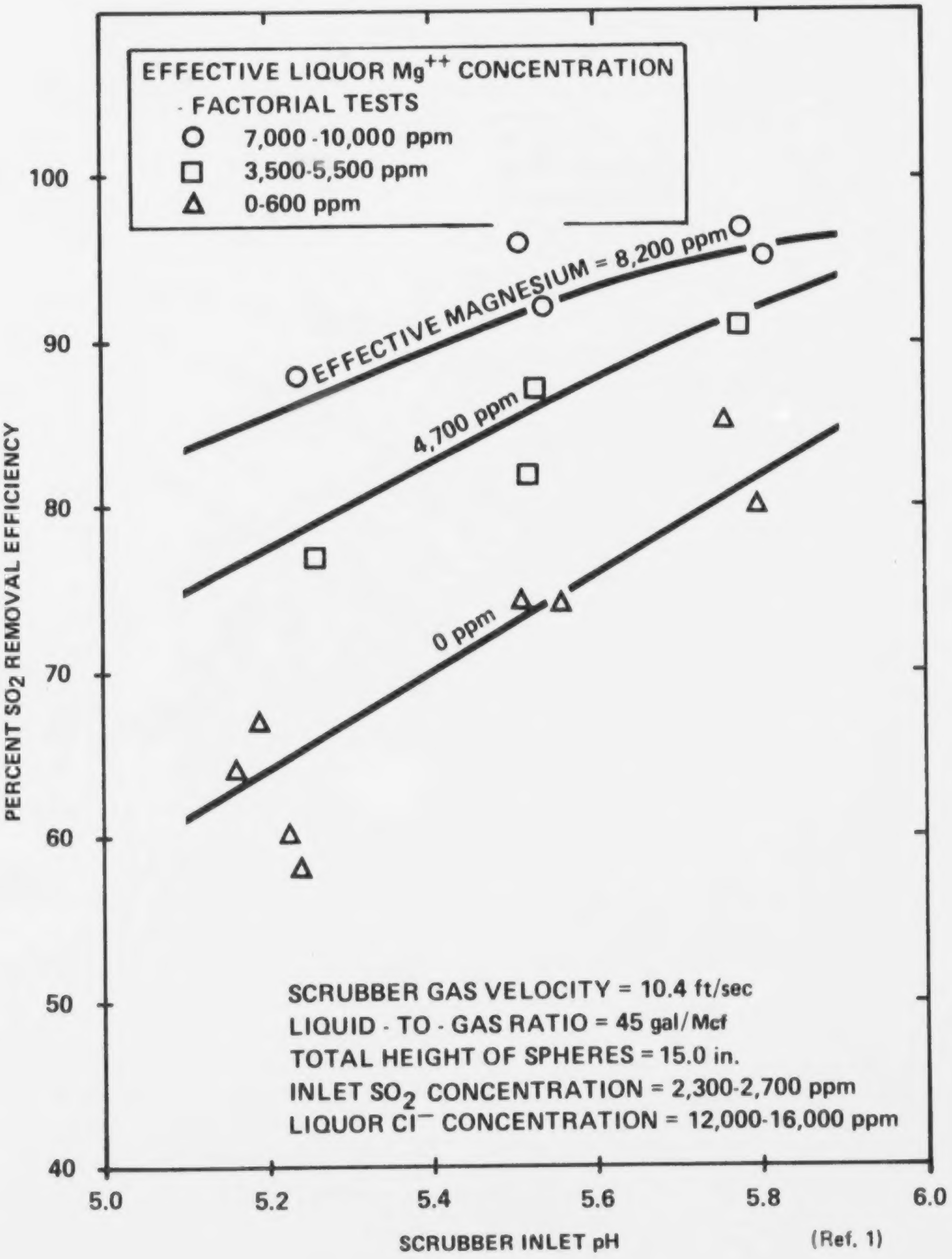


Figure 4-8. Effect of Scrubber Inlet pH on SO_2 Removal Efficiency — TCA with Limestone and Magnesium

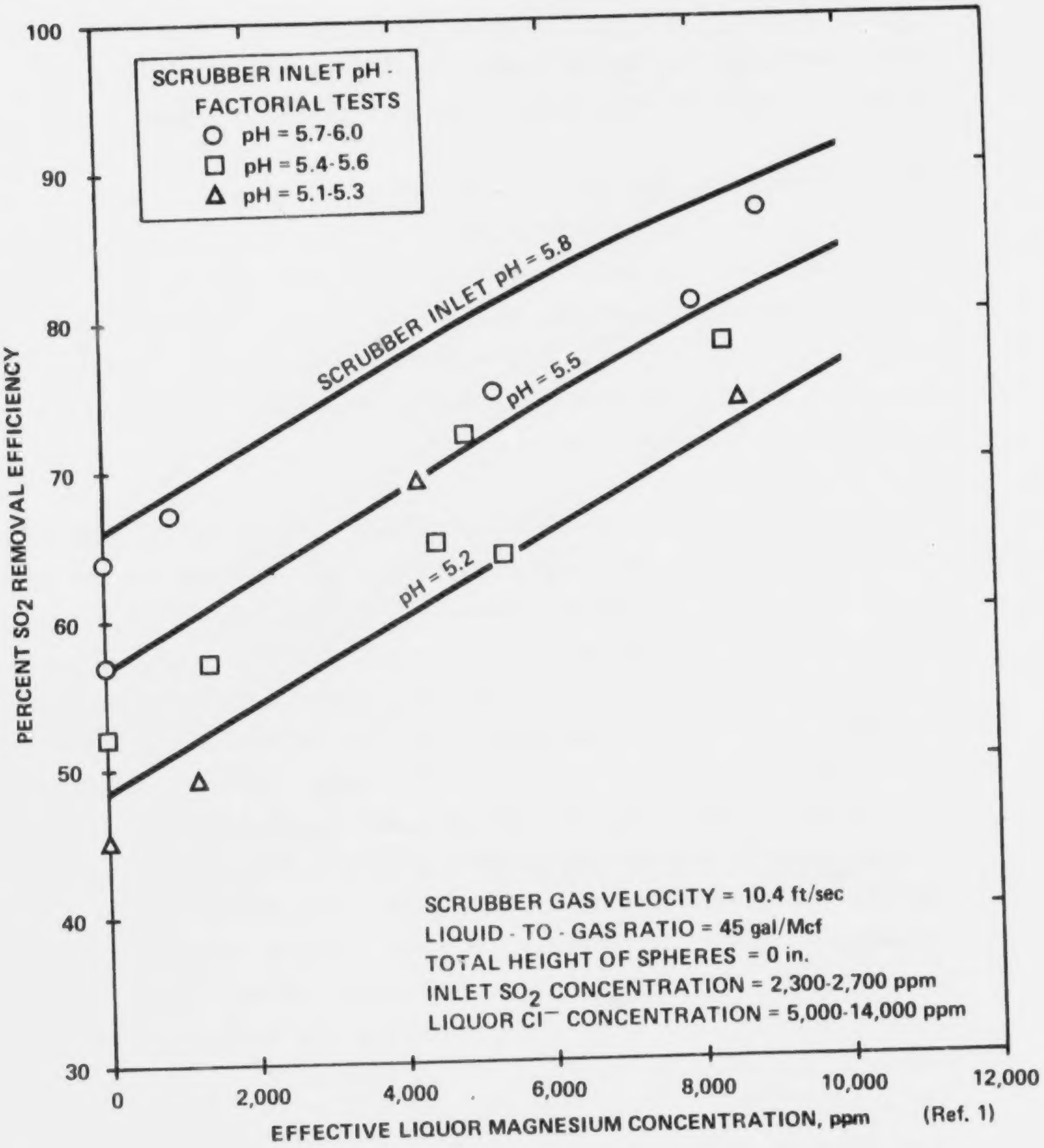


Figure 4-9. Effect of Magnesium on SO_2 Removal Efficiency — TCA (no spheres) with Limestone

Number of Scrubbing Stages. If a single scrubber does not provide sufficient SO_2 removal efficiency, it is possible, with increased cost, to design for two scrubbers in series. The net overall efficiency is substantially higher than that of either stage.

Figure 4-10 shows the variation in two-stage absorption efficiency as a function of individual scrubber efficiency. In this case the first scrubber is a relatively inefficient (for lime or limestone without additives) scrubber, such as a venturi. When the venturi and spray tower have been operated together at the Shawnee Test Facility (see Figure A-1), SO_2 removals greater than 90 percent have been realized during short-term tests.

When two scrubbers are operated in series, the mechanical complexity increases and additional energy penalties are incurred for the greater pumping requirements and gas pressure drop (see Figures 2-2 and 2-3). Japanese practice with oil combustion and recent U.S. developments in coal combustion indicate that the extra capital cost of two-stage scrubbing may be offset somewhat by the use of forced oxidation techniques to bring down the high costs of sludge disposal and improve alkali-utilization. In lime and limestone scrubbing, the waste product is normally a slurry of calcium sulfite, calcium sulfate (gypsum) and fly ash (if removed by the scrubber). The calcium sulfite can be oxidized to gypsum by air-slurry contact (forced oxidation). The resultant product has improved properties including higher settling rates, improved dewatering characteristics, and reduced total waste volume. One successful approach is to oxidize the slurry in the first of two scrubbing stages. The cost offset of two-stage forced oxidation does not apply to coals of lower sulfur content where oxidation and settled sludge density are normally high anyway. The relationship between sulfur content, oxidation, impounded sludge behavior, and cost is not yet fully quantified.

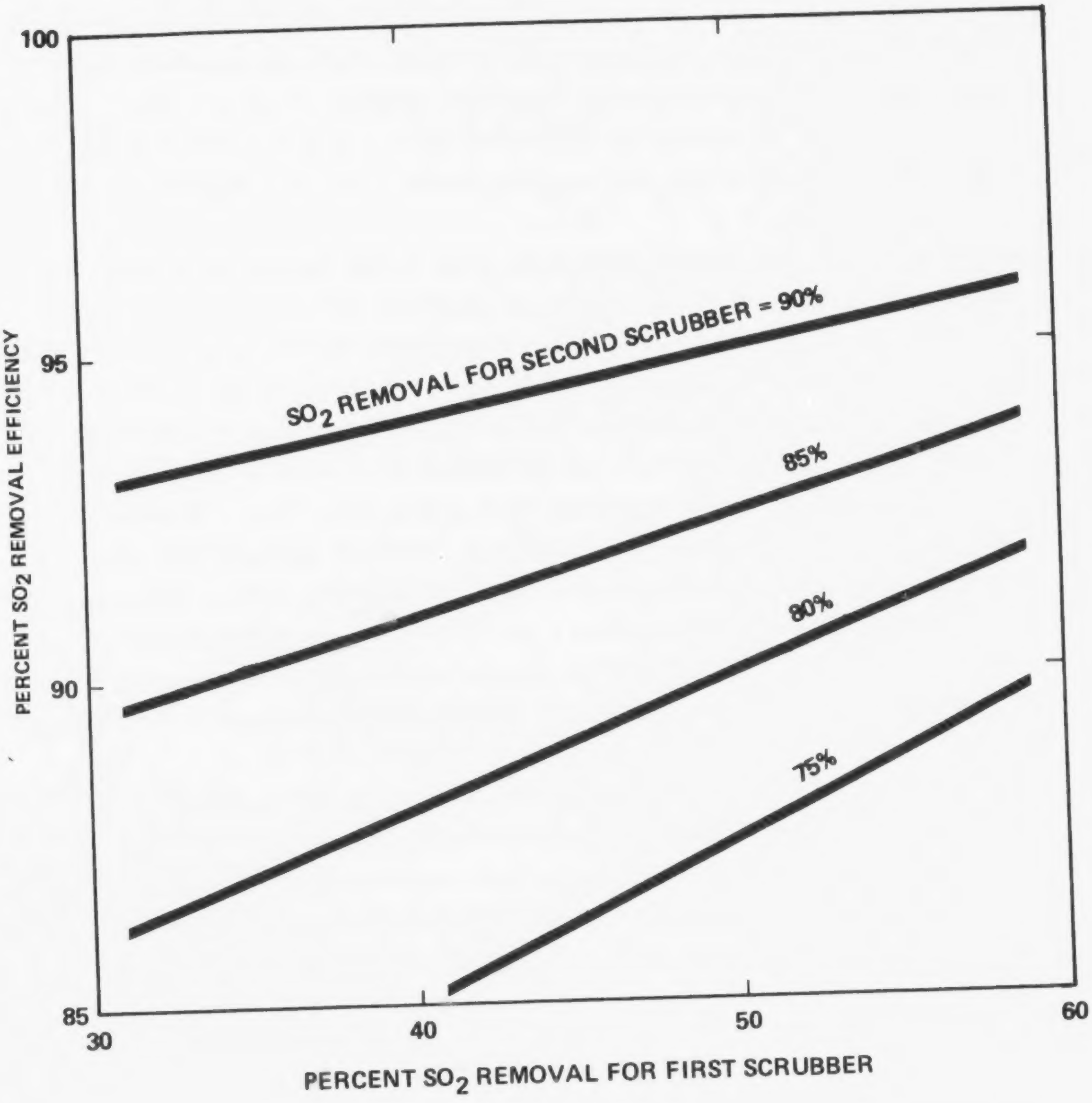


Figure 4-10. SO_2 Absorption Efficiency for Two Scrubbers in Series

Lime versus Limestone. The relative effects of limestone and lime scrubbing on efficiency in the Shawnee TCA, spray tower and venturi scrubbers at different pH's are shown in Figure 4-11 for selected run conditions. The performances of the three scrubber types are not directly comparable because of differing L/G's. However, the lime and limestone operating ranges for a given scrubber can be compared.

The fineness of the limestone particle size in the scrubbing slurry is largely determined by the degree of grinding before addition to the system. Lime, on the other hand, undergoes chemical reaction in the hold tank to precipitate very fine particles of calcium sulfite and calcium carbonate which then become the active alkali in the lime system. The limestone system also precipitates calcium sulfite, but the suspended alkali particles tend to be finer with lime. Because the particle size is larger for limestone, there is less surface area for dissolution so that, at equal stoichiometric feed ratios, the limestone exhibits slower dissolution rates than lime and, therefore, lower absorption efficiencies. To obtain similar SO_2 removals, the limestone system must be operated at higher stoichiometries.

In practice, when the limestone and lime processes are operated at maximum reliable pH's (pH's above which scaling may occur) of about 5.8 and 8.0 at the Shawnee facility, they achieve similar absorption efficiencies. In this case the limestone system would be operating at a feed stoichiometry of about 1.4 while the lime system would be at about 1.1.

In Figure 4-11, at a limestone pH of 5.8 and a lime pH of 8.0, the TCA shows almost identical efficiency, the spray tower shows slightly higher efficiency using lime, and the venturi shows better efficiency using limestone. For these same pH values, it is possible to come up with slightly different quantitative effects for limestone versus lime than shown in the figure by altering such variables as L/G, liquid holdup times, inlet SO_2 concentration, and feed stoichiometry.

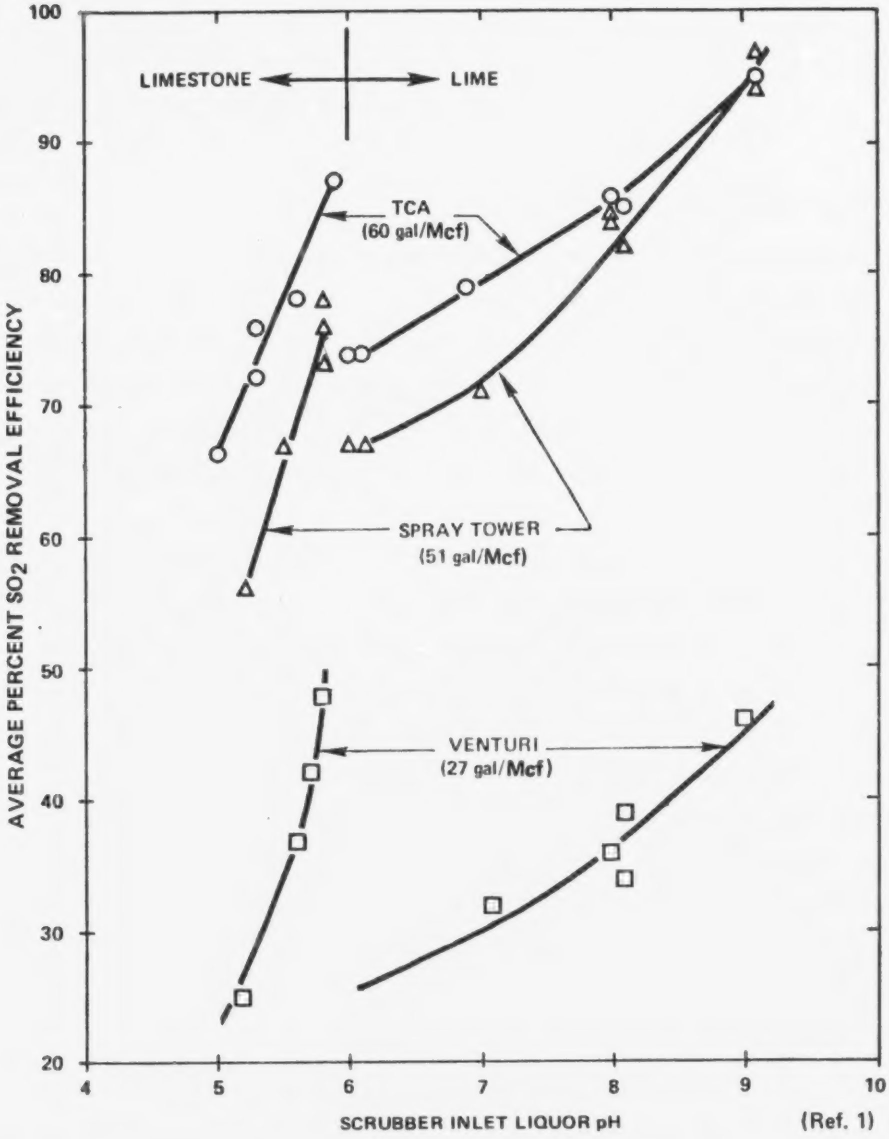


Figure 4-11. Effect of Scrubber Inlet Liquor pH on SO_2 Removal Efficiency — Lime and Limestone Derived Slurries

It should be noted again that the SO_2 equilibrium back pressures are very low, both when the limestone process slurry pH values range from 5 to 6 and when the lime process slurry pH values range from 7 to 8, so that the higher lime pH values do not hold an advantage in this respect.

Extrapolation of Data

Approximate equations and nomographs for predicting absorption efficiencies in limestone and lime wet scrubbing systems are presented in Appendix B. The equations are semitheoretical in form and have been fitted to spray tower and TCA SO_2 removal data from the Shawnee Test Facility (Ref.1). The nomographs cover approximately the same ranges of operation as the data to which the equations were fitted.

For a number of reasons the efficiencies predicted by these equations may not agree completely with that of a commercial installation. Among these are differences in scrubber internal configuration and liquid holdup, nozzle pressures, limestone quality and grind, lime quality, and variables that can affect liquor reactivity, such as sodium ion concentration, sulfate saturation, and sulfite oxidation. Nevertheless, the equations and figures in this section and Appendix A are significant in that they illustrate the relative effects of important process variables, and contribute to the technical data which must be available to permit design of full-scale units to meet specific local and regional conditions.

Effect of Higher Absorption Efficiency on Process Design

Raising the design efficiency of a lime or limestone FGD system from 65 or 85 percent to 90 percent or greater affects a number of process design parameters. Among possible required changes:

- Scrubbers. More elaborate and equally reliable scrubber modules must be engineered to provide greater slurry holdup and gas-liquid interfacial area, or existing design must be upgraded to include additional internal devices. The number of vessels could be doubled to provide operation with two scrubbers in series.

- Recycle Pumps and Piping. If L/G is increased to improve absorption efficiency, pump size is increased. Recycle slurry pipe diameters must be increased to maintain design flow velocities

- Limestone Preparation Equipment. To accommodate greater feed rates in a limestone system, larger grinding equipment, storage silos, transfer equipment, slurry makeup tanks and feed pumps, and other related equipment will be needed

- Lime Preparation Equipment. To accommodate greater feed rates in a lime system, the slaking equipment, storage silos, transfer equipment, slurry tanks and feed pumps, and other related equipment must be larger

- Magnesium Addition Equipment. If magnesium is added to improve efficiency, the magnesium handling equipment must be added to the process design, or dolomitic lime used. The waste sludge must be treated or contained to prevent groundwater contamination by the dissolved compounds

- Scrubber Recirculation Tanks. Tanks must be larger to maintain design residence time for increased L/G and to provide residence time for increased SO_2 absorption. These tanks may be directly under the scrubber, integrated into the scrubber vessel design, or offset. Slurry agitators will require redesign

- Slurry Solids Separation Equipment. This equipment must be larger to accommodate greater solids discharge. This applies to clarifiers, filters, centrifuges, and all related slurry pumps and piping

- Fans. If greater gas pressure drop is incurred, the fans must either be larger, or there must be more of them — with corresponding increases in the complexity of the fan control system

- Reheat. Bypass reheat would be marginal or impossible — because all the flue gas must be treated

- Nonair Quality Environmental Effects. As SO_2 removal is increased, provision must be made for handling the increased waste products produced. If soluble additives are used to enhance absorption efficiency, steps must be taken to prevent groundwater contamination

- Energy Requirement. Power requirements would increase with larger equipment loads

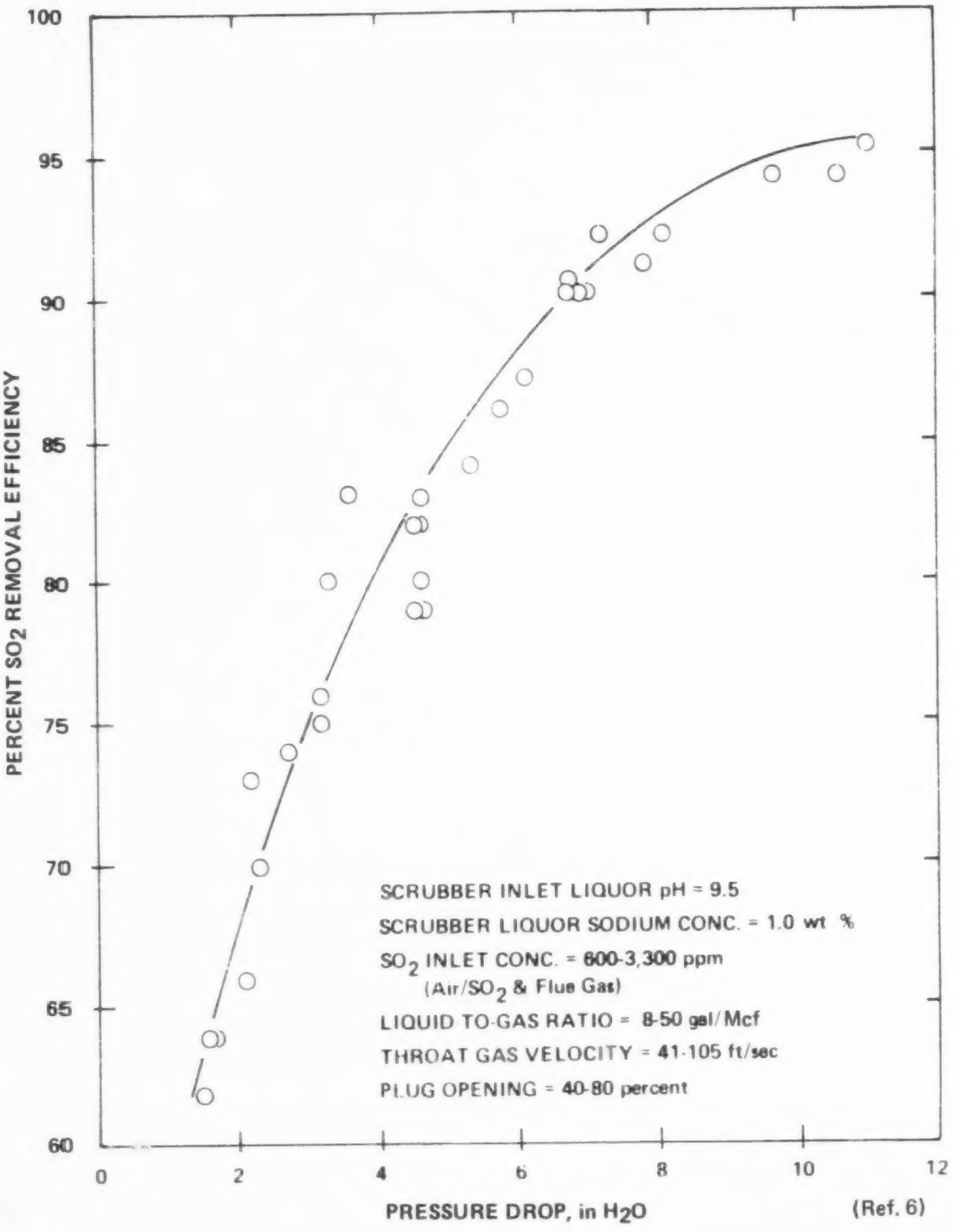
This list illustrates the complexity of considerations that must be addressed in attempting to design lime and limestone FGD systems for 90 percent or greater SO_2 removal efficiencies. Such an effort inevitably involves higher cost and energy penalties. The energy penalty will be in the form of increased auxiliary power consumption. The cost penalties can be not only in terms of heavier equipment, but of greater operating complexity and higher maintenance as well.

SODIUM CARBONATE SCRUBBING

The once-through sodium carbonate process is capable of high efficiency over a wide range of inlet SO_2 concentrations. This clear liquor process is gas-phase controlled — rather than being limited by the slow dissolution of suspended solids. However, the process consumes a premium chemical and produces a soluble waste salt which is normally discharged to a sealed (lined) evaporation pond for drying in arid areas.

During 1972 a series of tests was carried out at the Shawnee Test Facility using sodium carbonate (Ref. 6). Runs were made with air containing injected SO_2 and with flue gas from the boiler. The tests were designed primarily to determine coefficients in mathematical correlation models for predicting SO_2 removal efficiencies.

Figure 4-12 shows the effect of pressure drop on SO_2 absorption efficiency. Better than 90 percent efficiency was realized at a pressure drop of 15 mm Hg (8 in. H_2O) in these tests. The effects of L/G and gas velocity on efficiency in the TCA with no internals (operated as a spray tower) are shown in Figure 4-13. When the TCA was operated in its normal three-bed configuration, efficiencies higher than 99 percent were realized. Figure 4-14 shows the effects of L/G and marble



SCRUBBER INLET LIQUOR pH = 9.5

SCRUBBER LIQUOR SODIUM CONC. = 1.0 wt %

SO_2 INLET CONC. = 600-3,300 ppm
(Air/SO_2 & Flue Gas)

LIQUID TO-GAS RATIO = 8-50 gal/Mcf

THROAT GAS VELOCITY = 41-105 ft/sec

PLUG OPENING = 40-80 percent

(Ref. 6)

Figure 4-12. Effect of Pressure Drop on SO_2 Removal Efficiency — Venturi with Sodium Carbonate (10 MW size)

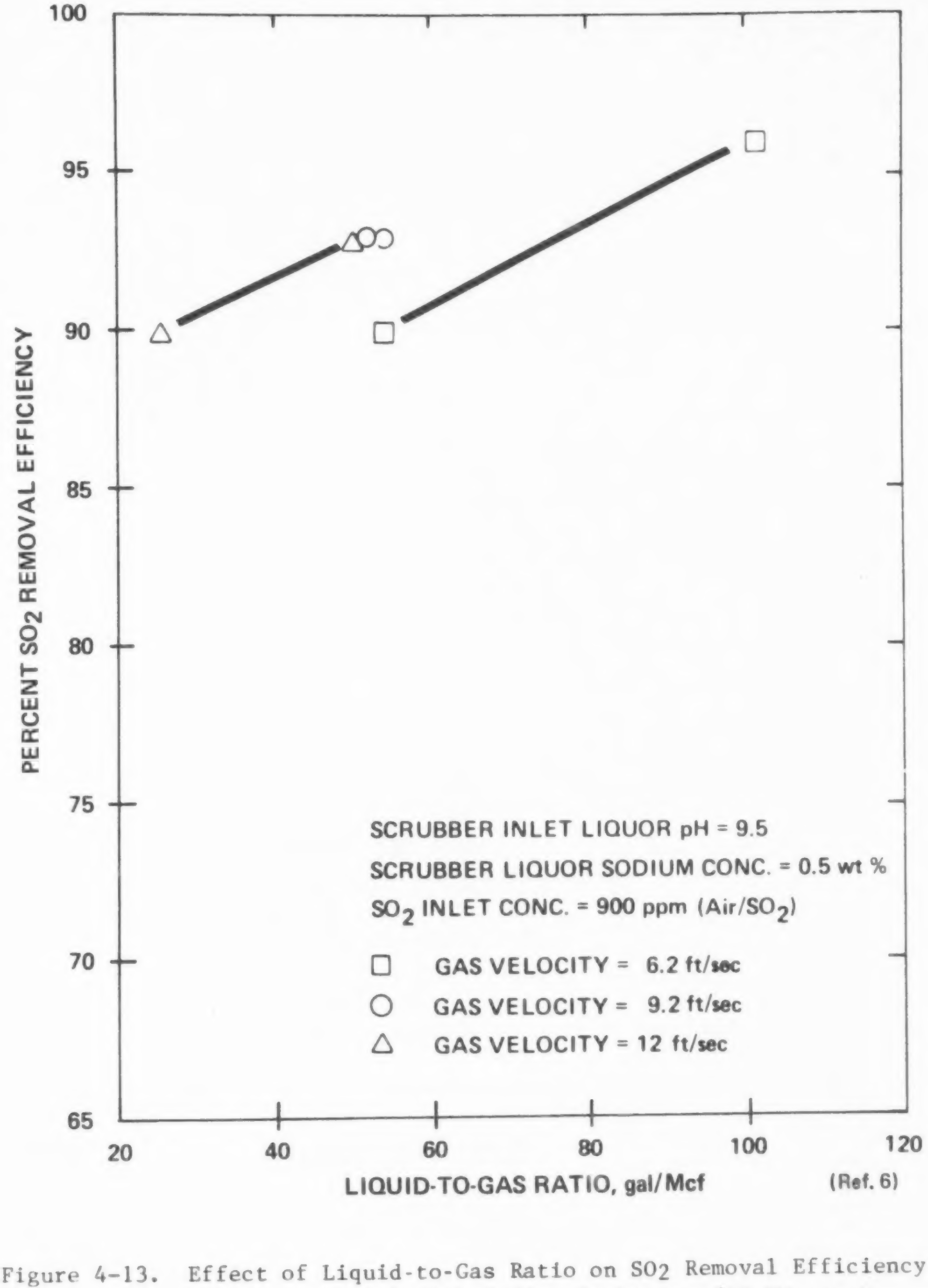


Figure 4-13. Effect of Liquid-to-Gas Ratio on SO_2 Removal Efficiency — TCA (no spheres) with Sodium Carbonate (10 MW size)

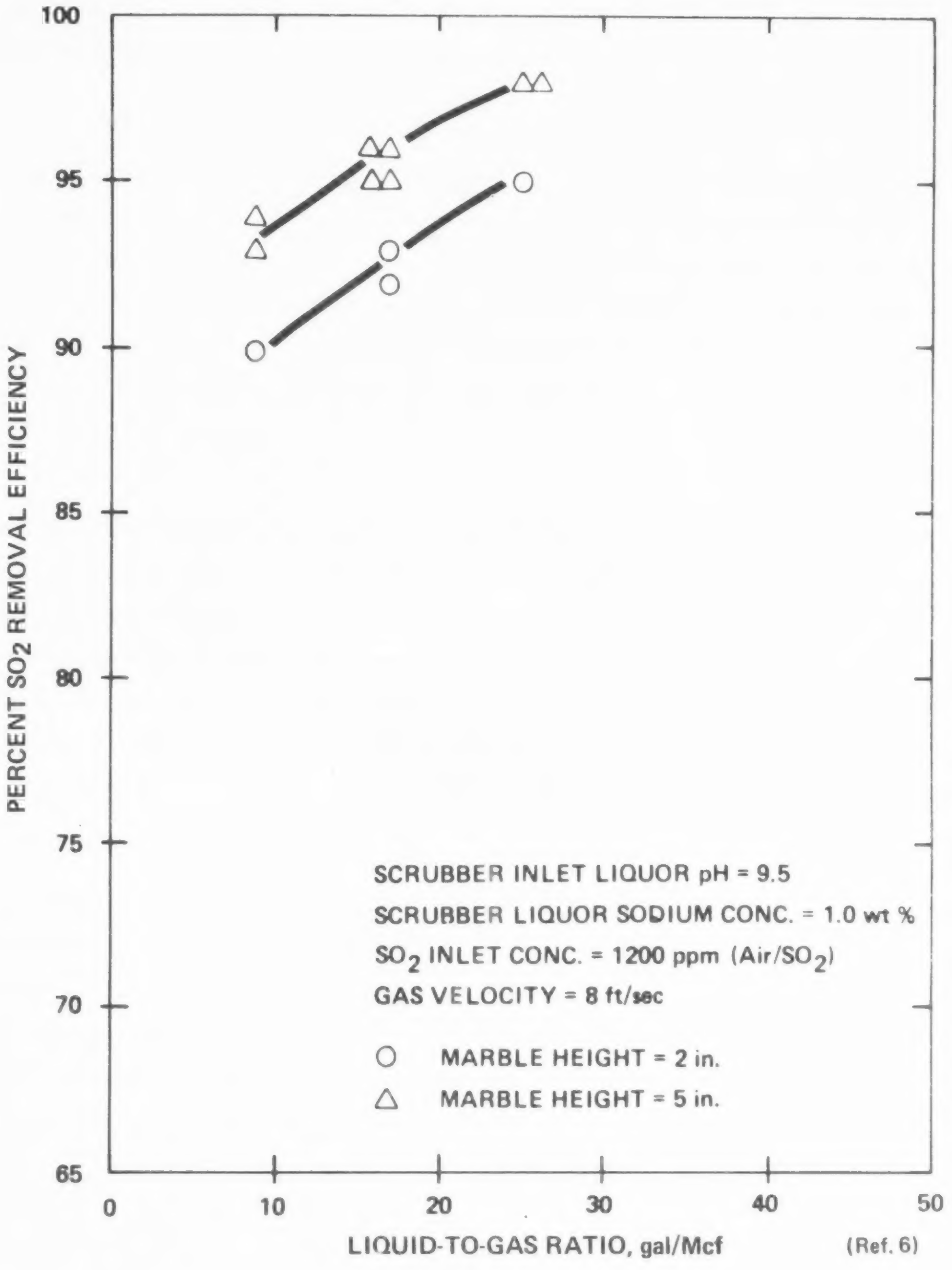


Figure 4-14. Effect of Liquid-to-Gas Ratio on SO_2 Removal Efficiency — Marble Bed Scrubber with Sodium Carbonate (10 MW size)

bed height on SO_2 removal in the marble bed scrubber. Absorption efficiencies of 98 percent were obtained for inlet SO_2 concentrations of 1,200 ppm.

To date application for coal-fired utility plant FGD systems has been limited to geographical areas having a cheap source of low grade carbonate. These include 3 units operating at a power station in Nevada (Reid Gardner - 375 MW) and one under construction in Wyoming (Jim Bridger - 520 MW). Both stations burn low-sulfur coal and the FGD systems are designed to realize about 90 percent removal efficiency.

DOUBLE ALKALI SCRUBBING

The clear-liquor sodium-based double alkali processes, like the once-through sodium carbonate, are aimed at realizing high removal efficiencies over a wide range of inlet concentrations. The process is gas-phase rather than solid dissolution limited. More efficient scrubber types can be used, such as packed towers and tray towers. Venturi scrubbers provide excellent SO_2 removal, although at substantial energy penalty. The process is more complex than the lime and limestone scrubbing processes.

Approximately 12 installations totalling 700 MW equivalent have been operated in oil-fired and coal-fired industrial boilers in the U.S. and Japan. Three installations totaling 1,050 MW on oil-fired utility boilers have operated in Japan, and 20 MW prototype testing has been carried out on a Gulf Power Company coal-fired utility boiler in Florida. The double alkali process remains to be tested at full-scale on a coal-fired boiler.

During 1973 an 8-month test program with a 3,400 m^3/hr (2,000 cfm) double alkali pilot plant was carried out by Combustion Equipment Associates and Arthur D. Little for the EPA to generate design data for the prototype system at the Sholz Steam Plant of Gulf Power Company.

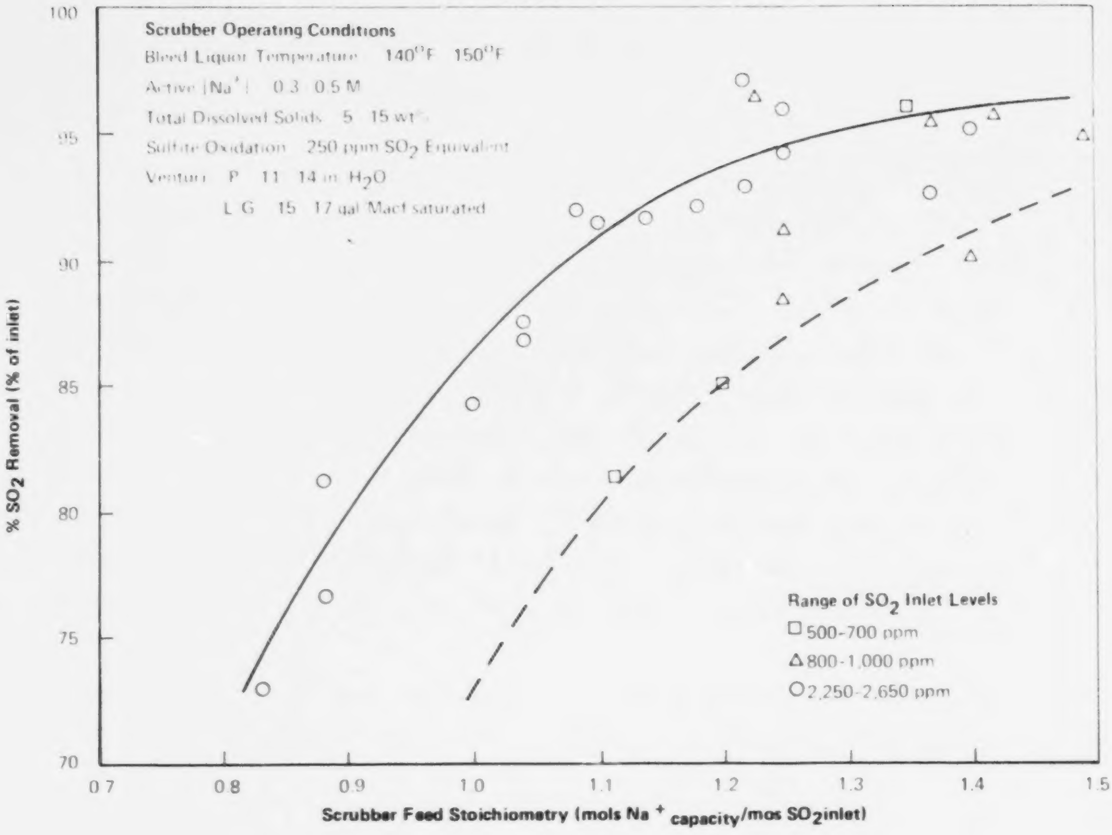
Results of the laboratory program and pilot operations are reported in the literature (Ref. 7). Figure 4-15 illustrates the high SO_2 removal efficiency possible with the double alkali processes.

The 20 MW prototype system was operated at the Sholz plant during 1975 and 1976. The scrubber system consisted of a venturi followed by a tray tower and was designed to remove a minimum of 90 percent of the SO_2 in the flue gas from medium to high sulfur coal (up to 5 percent sulfur, Ref. 7).

The system was operated with the venturi and absorber together in series and with the venturi alone. The absorption efficiency for each of these configurations is shown in Figure 4-16 for an inlet SO_2 range of 1,050 to 1,250 ppm. According to Reference 8, the points in the figure each represent at least a few hours of operation at the condition shown. Achieving a given outlet SO_2 level (within the efficiency limit of each configuration) was essentially a matter of adjusting the operation pH of the scrubber system. The scrubber L/Gs are not indicated in the figure. However, the 2.0 to 2.3 l/m^3 (15 to 17 gal/Mcf) L/G operation in Figure 4-15 shows that the L/G of a double alkali process is much lower than for lime and limestone processes.

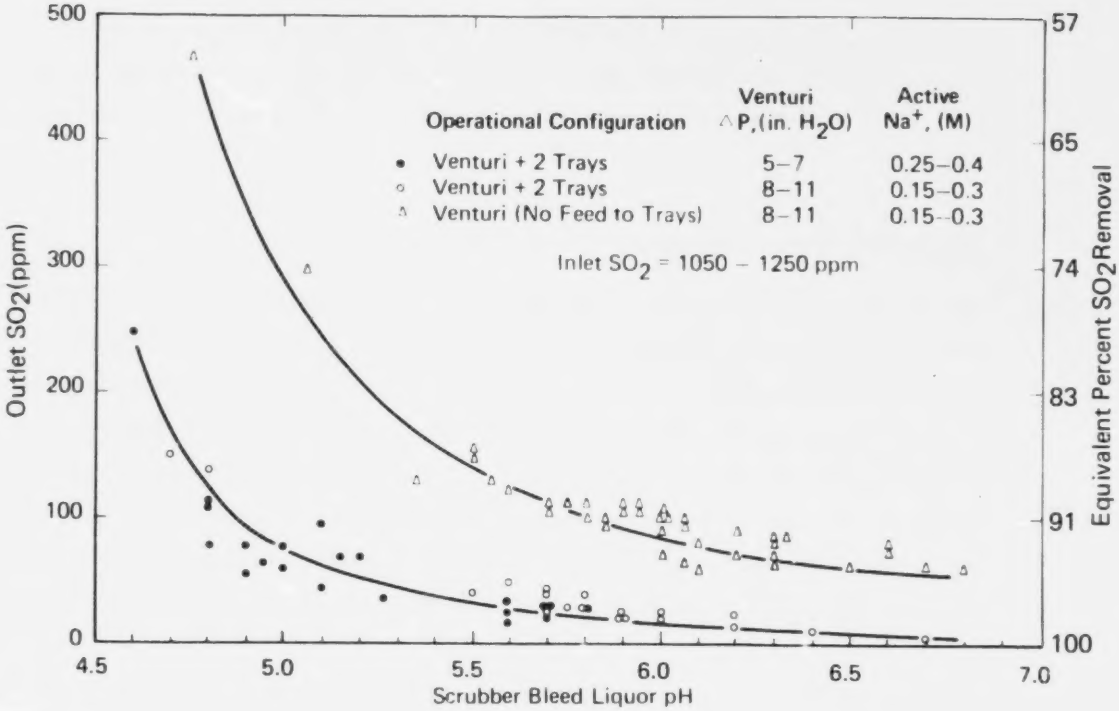
Using the venturi and tray tower in series, outlet SO_2 levels below 50 ppm were realized at a venturi liquor pH above 5.2, corresponding to better than 95 percent absorption efficiency. Above a pH of 6.0, outlet SO_2 levels dropped to 20 ppm or less - greater than 98 percent removal.

Generally, when the venturi and tray tower were operated together, the pH of the venturi bleed liquor was maintained between 4.8 and 5.9. Consequently, outlet SO_2 levels ranged between 15 and 100 ppm and, typically, between 30 to 70 ppm according to Reference 9. During one 2-month period the outlet SO_2 level ranged from 5 to 210 ppm and averaged 45 to 50 ppm. On five occasions the outlet SO_2 exceeded 100 ppm.



(Ref. 7)

Figure 4-15. Effect of Feed Stoichiometry on Removal Efficiency in the Venturi/2-Tray Tower Absorber for the EPA-ADL Double Alkali Pilot Program



(Ref. 8)

Figure 4-16. Effect of pH on SO_2 Removal for CEA/ADL Double Alkali Prototype System

Double alkali FGD systems are currently being offered and designed to achieve high absorption efficiency in both industrial and utility applications, although the process has yet to be demonstrated on a full scale coal-fired utility boiler. A 227 MW installation at Louisville Gas & Electric's Cane Run Station is scheduled to start up in 1979. The system is designed to meet performance guarantees of SO_2 emissions no greater than 200 ppm at coal sulfur contents up to 5 percent and 95 percent removal for coal containing over 5 percent sulfur.

MAGNESIUM OXIDE SCRUBBING

The magnesium oxide slurry scrubbing process has seen limited full scale operation in the U.S., although several systems operate in Japan on various industrial plant off-gases. A 155 MW installation was operated intermittently during 1972 through 1974 at Boston Edison's oil-fired Mystic Station. The system was designed for 90 percent SO_2 removal, but experienced mechanical operating difficulties. The system was generally operated at more than 90 percent removal throughout the program (Ref 10) and, during tests in 1974, averaged 91 percent removal.

A 95 MW magnesium oxide system was operated during 1974 and 1975 at Potomac Electric's coal-fired Dickerson Station. Although scrubber availability was not all that was desired due to problems with logistics in supplying raw materials (MgO), and to mechanical problems mainly attributed to under-design in such areas as piping, slurry pumps, and other auxiliary equipment, it was felt that the basic concept and design have been proven to be feasible (Ref. 10). SO_2 removal efficiency was reported to be consistently above 90 percent whenever operating at design gas flow and inlet SO_2 concentrations above 1,000 ppm (Ref. 11). During performance testing, removal efficiency ranged from 88 to 96 percent, depending on the gas flow rate (Ref. 12).

A 120 MW installation at Philadelphia Electric's coal-fired Eddystone Station was operated briefly in 1975, then shut down to move the MgO regeneration portion of the process to a nearby location. The unit started up again in 1977. The scrubber train consists of a venturi particulate scrubber followed by a venturi rod absorber. The system was designed to achieve better than 90 percent SO_2 removal from 2.5 percent sulfur coal (Ref. 13).

The magnesium oxide slurry scrubbing process has been demonstrated to be feasible on full-scale coal-fired boilers and is capable of SO_2 removal efficiency greater than 90 percent although the reliability of these installations must be improved. The process is more complex and costly (Refs. 14 and 15) than lime and limestone scrubbing processes.

SODIUM SULFITE (WELLMAN LORD) SCRUBBING

SO_2 removal efficiencies greater than 90 percent for this process are well established. About 17 installations are in operation in the U.S. and Japan on industrial plants and oil-fired boilers (Ref. 16). A 115 MW installation on the coal-fired Mitchell Station of Northern Indiana Public Service has been completed and a comprehensive one year demonstration program is underway. This is the first application of the process on a coal-fired boiler.

All other Wellman Lord installations operating in the U.S. are in refineries and display SO_2 removal efficiencies greater than 90 percent. Like other clear liquor scrubbing processes, the absorption efficiency is not limited by the slow dissolution of lime or limestone. The process is scale free and does not recirculate a slurry. Present scrubbers in this service are both efficient and reliable. The process is more complex and costly (Refs. 14 and 15) than the lime and limestone scrubbing processes.

The installation at Mitchell Station was designed to meet performance guarantees of a minimum of 90 percent SO_2 removal from flue gas generated by 3.5 percent sulfur coal (Refs. 17, 18) and has demonstrated greater than 90 percent removal.

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Section 5

EFFECT OF COAL PROPERTIES ON FGD SYSTEMS

The major coal properties affecting FGD system design and operation are heating value and sulfur, ash, moisture, and chlorine content.

The effects include:

- Heating value of coal. Affects flue gas flow rate — generally higher for lower heating value coals which also contribute a greater water vapor content to the flue gas
- Moisture content. Affects the heating value and contributes directly to the moisture content and volume of the flue gas
- Sulfur content. The sulfur content together with the allowable emission standards determines the required SO_2 absorption efficiency, the FGD system complexity and cost, and also affects sulfite oxidation
- Ash content. May affect FGD system chemistry and increases erosion. In some cases it may be desirable to remove fly ash upstream from the FGD system
- Chlorine content. May require high alloy metals or linings for some process equipment and could affect process chemistry or require prescrubbing

The importance of these factors is described in this section.

Coals are commonly classified according to rank, which refers to the degree of geological transformation from lignite to anthracite. A number of classification systems have been developed, but in very broad terms coal is ranked as anthracite, bituminous, subbituminous, and lignite. Typical ranges of coal properties for these categories are presented

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in Table 5-1. Bituminous coals are largely used by the U.S. power industry, but subbituminous coals and lignite are increasing in importance as supplies of high grade coal are depleted. Figure 5-1 indicates the geographical distribution of the various classes of coal deposits.

The properties of power plant flue gas depend on the complexities of the coal composition and on the equipment and conditions of the combustion process. An overview of the coal-flue gas relationship is possible by considering the apparent differences in the major combustion products from coals of widely differing rank fired in boilers appropriate for the efficient combustion of each. For example, Table 5-2 compares the heating values, ultimate analyses, wet flue gas analyses, firing rates, and SO_2 and fly ash emissions for a bituminous coal fired in a pulverized coal boiler and for a lignite fired in a cyclone boiler, both at 130% total air.

The above comparison should not be regarded as necessarily typical because similarly-ranked coals may vary widely in composition.

The data show that the coals are significantly different with respect to heating values (related to their differing ratios of carbon to oxygen) and their sulfur, moisture, ash, nitrogen, and chlorine content. The major significant differences between the combustion gas streams is the volume per megawatt, moisture content, SO_2 and fly ash concentrations, and to a lesser degree, NO_x and chloride content. In general, larger flue gas volumes per megawatt result from coals with lower heating values.

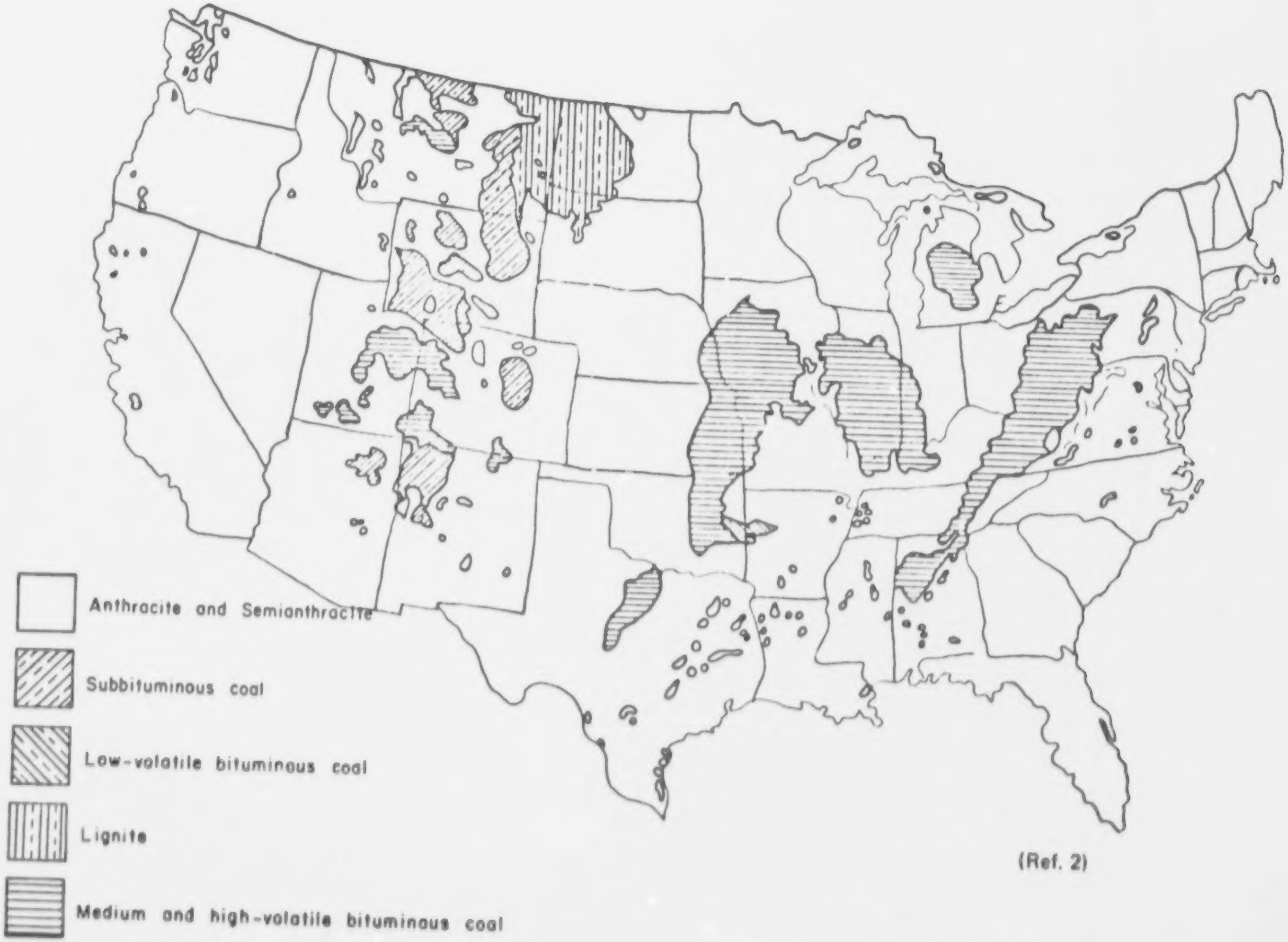
The following sections summarize the major effects of coal and its combustion gas characteristics on FGD processes.

Table 5-1

TYPICAL RANGE OF COAL ANALYSES

	Anthracite	Bituminous	Subbituminous	Lignite
Analysis:				
Moisture, %	2-5	2-15	15-30	25-45
Volatile matter, %	5-12	18-40	30-40	25-30
Fixed carbon, %	70-90	40-75	35-45	20-30
Ash, %	8-20	3-25	3-25	5-30
Heating value, 10^3 Btu/lb	12-14.5	10-14	8-10.5	5.5-8
Sulfur, %	< 1	0.5-5	0.5-3	0.5-2.5
Nitrogen, %	0.5-1	1-2	1-1.5	0.5-1.5

(Ref. 1)


Anthracite and Semianthracite
Subbituminous coal
Low-volatile bituminous coal
Lignite
Medium and high-volatile bituminous coal

(Ref. 2)

Figure 5-1. Coal Fields of the United States

Table 5-2

EXAMPLES OF COAL AND FLUE GAS COMPOSITIONS

Characteristic		Bituminous	Lignite
Coal	Heating value, Btu/lb	11,700	7,070
	Carbon	64.2 wt %	42.6 wt %
	Hydrogen	4.4	2.9
	Oxygen	6.5	11.8
	Nitrogen	1.3	0.6
	Sulfur	4.0	0.7
	Chlorine	0.1	—
	Ash	11.5	6.4
	Moisture	8.0	35.0
		100.0	100.0
Flue Gas	Nitrogen	74.29 vol %	69.06 vol %
	Carbon dioxide	12.66	12.67
	Oxygen	4.52	4.20
	Water vapor	8.17	13.91
	Sulfur dioxide	0.30	0.08
	Sulfur trioxide	0.003	0.0008
	Nitrogen oxides	0.05	0.07
	Hydrogen chloride	0.007	—
		100.00	100.00
	SO_2 concentration, ppm	3,000	800
	Fly ash, grain/scf (wet)	4.3	1.3
	Coal firing rate, lb/hr/MW	795	1,460
	Gas rate, acfm/MW @ 300°F, 1 atm	3,100	3,760
	SO_2 emission: lb/hr/MW	64	20
	lb/10^6 Btu	6.9	1.9
	Fly ash emission, lb/hr/MW	78	28
	lb/10^6 Btu	8.4	2.7

Note: Similarly-ranked coals vary so widely in composition that these examples should not be regarded as typical.

Basis:

30% excess air; 100% sulfur overhead
Bituminous coal: pulverized coal boiler; 9,300 Btu/kWhr boiler heat rate; 85% ash overhead
Lignite coal: cyclone boiler; 10,300 Btu/kWhr boiler heat rate; 30% ash overhead

COAL HEATING VALUE AND MOISTURE CONTENT

Coal Firing Rate

Because a power plant using a low heating value coal must fire at a higher burn rate to generate the same amount of power, such coals produce a larger volume of flue gas and greater SO_2 emissions per unit of generated power. The effect on the FGD system is twofold. First, the flue gas handling equipment, including the scrubbers, must be of a larger size to accommodate the greater gas volumes. Typically, power plant flue gas volumes may range from 5,000 to more than 7,000 m^3/hr/MW (about 3,000 to 4,000 acfm/MW), depending on the coal composition, boiler heat rate, gas temperature, and power plant elevation (or gas pressure). Secondly, the increased SO_2 emissions mean that on a megawatt basis the FGD system must treat proportionally larger quantities of SO_2. This is illustrated with the examples presented in Table 5-2 above. Here the higher heating value bituminous coal contained nearly 6 times as much sulfur as the lignite (4.0% versus 0.7%). Yet on a megawatt basis the bituminous coal SO_2 emissions were only 3 times as large (20 versus 9 kg/hr/MW, or 64 versus 20 lb/hr/MW).

On a megawatt basis, therefore, the FGD system (as well as the power plant) equipment capacity is greater and capital and operating costs are higher for coal with lower heating value — for a given coal sulfur content and SO_2 absorption efficiency (Ref. 3).

Water Vapor

A characteristic of lower heating value coals and coals of high moisture content is a flue gas with a greater proportion of water vapor. In a wet scrubber this means that less water is evaporated per unit volume of gas and the adiabatic saturation temperature for the gas (the temperature to which the gas is cooled in the scrubber by contacting the scrubbing liquor) is higher. This also results in higher recirculating slurry temperatures. The overall effect is that the SO_2 absorption process may take place at somewhat higher tempertures (perhaps 6 to 8°C, or 10 to 15°F, higher for lignite over bituminous coals). For the example in Table 5-2 the bituminous coal flue gas has a saturation temperature of 52°C (126°F), while the corresponding lignite gas saturation temperature is 59°C (138°F). Elevated temperatures could affect absorption efficiency depending on the chemistry of the particular process.

SULFUR CONTENT OF COAL

Sulfur is present in coal in amounts ranging from traces to 5 percent or more. Three forms of sulfur exist in the coal:

- Organic sulfur which is combined with the coal substance
- Pyritic sulfur which is combined with iron
- Sulfate in the form of calcium or iron sulfate

Sulfate sulfur accounts for less than one percent of the total sulfur present, while the proportions of organic and pyritic sulfur vary widely depending on the coal. Upon combustion most of the sulfur appears in the flue gas as sulfur dioxide. The inorganic sulfur content of coal can be reduced by washing, but no system has yet been commercialized for this purpose, and not all coals are amenable to such treatment.

SO_2 Removal Rate

The sulfur content of the coal together with the allowable emission standards determines the absolute removal rate of SO_2 in pounds per hour. For a given absorption efficiency the sulfur content of the coal directly affects the design of almost every piece of equipment in the FGD system.

For example, a lime or limestone system designed for high rather than low sulfur coal would have:

- Scrubbers with capacity for greater SO_2 removal
- Higher L/G's and therefore bigger pumps and piping and higher pumping energy requirements
- Bigger fans and greater energy requirements if the improved scrubber design results in higher gas pressure drop
- Larger sized alkali storage, preparation and feed equipment
- Greater lime or limestone feed rates
- Larger scrubber recirculation tanks to maintain residence time for increased L/G's and to provide additional time for increased SO_2 absorption load
- Greater capacity slurry solids separation equipment
- Provision for disposing of the larger waste volumes
- Increased power requirements for the larger equipment loads

The result is that for higher sulfur coals the lime and limestone FGD systems are more complex and have higher capital and operating costs (Refs. 3, 4).

Other FGD processes are similarly affected by the sulfur content of the coal. For systems using regenerable absorbents (double alkali, magnesium oxide and Wellman Lord processes), the capacity of the regeneration section is directly proportional to the sulfur content. With high sulfur coal the overall cost of these sections (capital and operating) represents a large portion of the total cost for the system. With recovery processes there is relatively little waste but a large by-product processing cost, although the greater amount of by-product produced helps to offset these costs.

Sulfite Oxidation

The amount of sulfite oxidized to sulfate in scrubber solutions is proportional to the relative amounts of oxygen and SO_2 absorbed. It also depends on the pH and temperature of the liquid as well as the composition of particulate emissions which may contain iron or copper that act as catalysts for the oxidation reaction. In general, however, as the gas SO_2 concentration becomes smaller, the fraction of sulfite oxidation tends to increase. For this reason, when lime or limestone scrubbing systems are used for low sulfur coal or for boilers operating on high excess air they may experience high sulfite oxidation and produce waste solids that are mainly gypsum. This can be a desirable feature since gypsum solids are more easily dewatered due to faster settling rates and higher final settled densities. Conversely the lower oxidation observed with high sulfur coals can lead to solid wastes high in sulfite and difficult to dewater.

In the double alkali process the formation of sodium sulfate by oxidation is normally undesirable since in concentrated solution in the presence of sodium sulfite it is relatively difficult to convert back to an active form of sodium. For high sulfur coal applications in which oxidation may be expected to be relatively low, a concentrated double alkali system can be employed. With this mode of operation (see Section 3) oxidation is controlled by using a concentrated absorbing solution which minimizes absorption of oxygen from the gas. For low sulfur applications where higher oxidation is unavoidable a dilute mode of operation is required. In the dilute solution the sodium sulfate reacts well with lime to precipitate gypsum and regenerate active sodium for recycle to the absorber. Thus a high degree of oxidation is not detrimental and forced oxidation is sometimes used to convert all the sulfite to sulfate.

Sulfur Trioxide

About one percent of the sulfur in the flue gas is present as sulfur trioxide (SO_3). When the gas is cooled below about 135°C (275°F), sulfur trioxide reacts with water vapor to form sulfuric acid. This acid condenses at temperatures well above the water dew point. If the acid droplets form in an atmosphere of water vapor, a fine mist is produced which is difficult to remove, even by wet scrubbing. If the reaction occurs by absorption of gas into the scrubbing liquor, the SO_3 is removed and contributes to sulfate formation.

ASH CONTENT OF COAL

Most coals fired in U.S. utility boilers contain 5 to 30 percent ash. After combustion, part of the ash falls to the bottom of the furnace and the remainder is carried upward with the flue gas. The fraction of the ash that is carried overhead is a function of the boiler design and combustion parameters. With pulverized coal firing, 85 percent or more of the ash appears as fly ash; in cyclone boilers about 20 to 30 percent of the ash goes overhead (Ref. 1).

Mineral Content

The principal inorganic constituents of fly ash are silicon, aluminum, iron, and calcium. The oxides of these four elements comprise 95 to 99 percent of the composition of ash. Ash also contains smaller amounts (0.5 to 3.5 percent) of magnesium, sulfur, sodium, and potassium as well as very small quantities (parts per million) of from twenty to fifty elements (Ref. 2). When fly ash containing potentially hazardous heavy metals is removed together with SO_2 the waste product may require special disposal precautions to prevent contamination of groundwater.

Fly Ash Removal

Fly ash can be removed upstream of the FGD system by a precipitator, fabric filter, or prescrubber, or integrally within the FGD system itself.

Not all FGD processes are suitable for combined removal. Even when fly ash is removed upstream, residual ash becomes entrained in the process liquor.

Fly ash is invariably abrasive; some is chemically inert, and some is highly acidic due to SO_3 adsorption. Fly ash can cause excessive erosion, scaling and plugging of equipment. It contributes to the waste volume of throwaway processes, the loss of absorbent for regenerative processes, and may contaminate the byproduct of recovery processes. Certain coals from Wyoming, Montana, and North Dakota produce alkaline fly ash with large amounts of reactive calcium, magnesium, sodium, and potassium oxides. With combined removal of alkaline fly ash and SO_2, major reduction in the alkali makeup requirement can be realized. The presence of calcium alkali in the ash can, however, aggravate wet dry interface problems by producing hard insoluble deposits.

In general, only processes using nonregenerable absorbents (lime, limestone, soda) can be used for the combined removal of fly ash and SO_2. The fly ash is then disposed of together with the spent absorbent. Lime and limestone processes can better tolerate this fly ash and are sometimes used for combined removal. Venturi scrubbers are often used for this purpose at the expense of increased pressure drop over other absorption systems. However, the fly ash contributes to solids buildup at the wet-dry interface and causes erosion of pipes, pumps, spray nozzles, and scrubber internals.

Soluble clear liquor soda processes can be designed for combined removal, but the FGD systems must then be designed for slurry handling and must deal with special disposal considerations. For combined removal with a soda system the fly ash must be free of calcium to retain scale-free operation of such systems.

Processes using regenerable absorbents (double alkali, magnesium oxide, and Wellman Lord) are not suited for combined removal of fly ash and SO_2.

A double alkali mode of operation, while theoretically feasible for joint removal, would suffer from high cost and absorbent loss. Fly ash could in some instances introduce a scaling problem. Similarly, the Wellman Lord system would become complex, absorbent loss would increase, and the system would produce a fly ash sludge contaminated by sodium sulfate. The magnesium oxide process cannot be used for combined removal because of contamination of the spent absorbent by the fly ash.

Alkaline Fly Ash Scrubbing

A significant characteristic of many western U.S. low sulfur coal ashes is their high content of calcium oxide and magnesium oxide. The alkali content tends to be highest in lignite and less prevalent in subbitumous and bituminous coals. In lignite, calcium oxide concentrations may be 20 percent or higher in the fly ash, while magnesium and sodium oxides may range up to 6 percent or more. These fly ash alkali values have been successfully used for the absorption of SO_2 with a major reduction in the raw alkali makeup requirement (Ref. 5).

The alkali content may be sufficient to react with and absorb a large portion of the SO_2 produced by the coal combustion. The fly ash can either be captured by the FGD scrubbers and recirculated with the absorption liquor, or it can be collected dry in a precipitator and then added to the process. A key design factor for alkaline fly ash scrubbing is that the alkalinity is generally released and usable only under low pH absorption conditions (Ref. 6). This can limit absorption efficiencies, but also tends to promote high oxidation of the waste product for improved disposal characteristics. Due to the natural variability of fly ash quantity and alkali content a supplemental feed of limestone or lime is generally required for effective chemical control of the system.

CHLORINE CONTENT OF COAL

Chlorine varies from a trace to as high as 0.5 percent in U.S. coals. It is rare in western coal, but tends to be common in the East. During combustion the chlorine is converted to hydrogen chloride or other volatile chlorides. Nearly all of the hydrogen chloride in flue gas is absorbed by the scrubbing liquor in a wet FGD system. Chloride concentrations in the gas may be as high as 100 ppm for coal with high chlorine content. Upon absorption, chloride concentrations build up in the recirculating scrubber liquor. The concentration is a function of the water loop closure of the FGD system, and levels may reach 5,000 to 10,000 ppm or more for high chloride coals and water loop closures corresponding to the discharge of a waste stream containing 60 percent solids or higher.

Stress Corrosion

The presence of chloride in the liquid circuits of an FGD system provides the potential for chloride stress-corrosion in alloys susceptible to the phenomena. The potential is related to concentration, temperature, pH, alloy treatment and stress levels. The solution to the problem requires the use of high alloy equipment wherever rubber or other protective coatings are not feasible.

Effect on Process Chemistry

In wet scrubbing processes, dissolved chloride replaces active calcium, magnesium or sodium alkalis by their chloride salts, which are inactive in the absorption process. The alkali associated with the chloride is then lost as dissolved solids in the water portion of the waste sludge. From a cost standpoint this is particularly objectionable for magnesium and sodium based processes (or magnesium enhanced lime and limestone processes), because these alkalis are relatively expensive. For such processes, prescrubbing may be used to absorb chlorides from the flue gas upstream of the FGD system. This minimizes both alkali loss and chloride stress-corrosion problems. For lime and limestone processes, an equivalent amount of calcium is used up by the chloride (the amount is small relative to that used for SO_2 absorption), but the calcium is relatively inexpensive.

NITROGEN CONTENT OF COAL

Nitrogen oxides are formed by oxidation of nitrogen compounds in the coal and by reaction of nitrogen and oxygen in the combustion air at high temperature. In utility boilers nitrogen oxides are typically about 95 percent nitric oxide (NO) and about 5 percent nitrogen dioxide (NO_2). Less than 0.1 percent of the nitrogen entering the boiler leaves as NO_x. Cyclone boilers operate at higher flame temperatures than other boilers, generating somewhat higher concentrations (Ref. 1).

Although nitrogen oxides are present in small concentrations (500 ppm or less in pulverized coal boilers), they may participate in SO_2 oxidation phenomena and thereby affect the chemistry of FGD systems (Ref. 7). Only small amounts of these oxides are absorbed, showing up as nitrates in the scrubber liquor. In most installations to date the level of nitrates has not been high and nitrogen oxides are normally not considered in designing a wet scrubbing FGD system.

FGD systems currently installed in the U.S. cannot effectively remove nitrogen oxides. Control of their ground level concentrations often establishes stack height and may contribute to requirements for reheat.

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Section 6

REHEAT OF SCRUBBED FLUE GASES

The reasons and methods for reheating scrubbed flue gases are described in this section. The energy consumption of the various methods is defined, and problems anticipated or experienced with reheater design, operation, and maintenance are described. Alternative approaches to reheat are discussed.

Flue gases are normally discharged from the regenerative air heater of a power plant at 120 to 150°C (250 to 300°F). The temperature is selected to remain above the dew point of the traces of H_2SO_4 normally present in order to reduce corrosion and permit carbon steel to be used for fans, ducting, and stack lining.

When a wet scrubber is inserted between the air heater and stack for SO_2 removal, the flue gas exiting the scrubber is saturated with water and cooled to the saturation temperature of about 50°C (125°F). Discharge of the cool, wet gas to the stack produces:

- Water condensation in the stack exit gas (acidic rain or fallout from the plume) and possibly in the exit duct and stack

- Duct work and stack corrosion precluding use of carbon steel

- Impaired plume rise and hence poorer dispersion for a particular stack height

To correct the above undesirable aspects of wet scrubbing, the treated gas may be reheated to a higher temperature before discharge.

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Although considerable attention has been focused on the importance of proper scrubber design, there is perhaps less understanding of the reheat function; yet, failure of a reheater can shut down the entire system. Investment and operating costs for reheating can be high. For these reasons the EPA has initiated a comprehensive reheat assessment study.

PURPOSE AND NEED FOR REHEAT

The reheater is basically a heat transfer and/or gas mixing device for heating the flue gas and restoring buoyancy by the addition of sensible heat. This sensible heat partially compensates for the latent heat losses that occur during adiabatic saturation within the scrubbers. The amount of energy added and the final flue gas temperature depend on the specific need for which the temperature increase is being considered.

A number of installations are operating under "wet stack" conditions, i.e., without reheat (see page 6-19). Such conditions approach the ideal design goal of minimizing the thermal energy penalties associated with flue gas cleaning. These installations provide substantial sulfur dioxide removal, and evidence to justify reheat is scarce. Some of the factors that could influence such evidence are discussed in the following paragraphs.

Ground Level Concentrations of Pollutants

Scrubbers are installed to reduce emissions and thereby reduce the exposure to pollutants at ground level. A stack provides dispersion of the plume so that concentration of pollutants will fall within accepted standards by the time any part of the flue gas reaches the ground.

An important aspect of this dispersion of the flue gas plume is the height to which the plume rises after it leaves the stack. The plume rise varies with the meteorological conditions, but is particularly dependent on the temperature of the gas. The higher the plume rise, the lower will be the peak ground-level concentrations downwind of the stack.

Without reheat the oxides of nitrogen, which are not significantly removed by the scrubber, may reach ground-level concentrations exceeding those which would have prevailed had no scrubbing occurred. Sufficient plume rise must be provided to prevent this.

When the design efficiencies for sulfur dioxide removal are realized, it is unlikely that the ground-level concentration downwind of the stack will be higher than without scrubbing.*

Duct and Stack Corrosion

One of the features which limits boiler efficiency is the effect of corrosion on heat recovery hardware as the temperature of the exit flue gas approaches the critical region of 120 to 150°C (250 to 300°F). Further heat recovery by cooling in a typical power plant requires corrosion-resistant construction, the cost of which cannot be justified.

During combustion, while most of the sulfur is converted to sulfur dioxide, a portion reacts further to form sulfur trioxide (SO_3). Within the scrubber these acid gases react to form sulfurous and sulfuric acids which are neutralized by the alkali. Beyond the scrubber outlet, water vapor continues to condense, reacting with residual SO_2 to form sulfurous acid. In the absence of alkali, an acidic, corrosive condition prevails.

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*Such a situation was reported in England at Battersea. Under certain conditions, the plume — which was not reheated — fell to ground immediately after leaving the stack, causing complaints by nearby residents. Battersea, however, employed once-through scrubbing and it seems probable that outlet gas temperatures fell well below the adiabatic saturation temperature characteristic of present day closed-loop scrubbers, thereby contributing to the severe plume drop.

Sulfur trioxide has a far more corrosive effect than sulfur dioxide. Concentrations may range from 0.5 to 5 percent of the total sulfur oxides present (Ref. 1). Even this low concentration of SO_3 raises the acid dewpoint temperature considerably. Condensation occurs, and corrosive sulfuric acid is formed.

Only partial removal of SO_3 occurs in the scrubber. During tests with a venturi scrubber at the EPA Shawnee Test Facility, inlet SO_3 concentration ranged from zero to 22 ppm, while emission concentrations ranged from zero to 14 ppm. Average SO_3 removal was about 60 percent (Ref. 2). For SO_3 concentrations in the flue gas between 6 and 8 ppm, theoretical acid dewpoints are between 127 and 132°C, or 260 and 270°F (Ref. 3).

The water dewpoint of scrubbed flue gas is typically about 50°C (125°F). The presence of as little as 0.1 ppm of SO_3 in the flue gas creates an acid dewpoint of above 93°C, or 200°F (Ref. 3).

It is questionable whether the reheat of such gas to the acid dewpoint can be justified. This may be further quantified by the EPA reheat assessment study. The very slight condensation of SO_3 with water depletes the acidity within the gas, thereby dropping the acid dewpoint downstream. In the final analysis, it is only necessary to reheat sufficiently to prevent significant condensation of corrosive vapor. This means heating the gas to a point intermediate between water dewpoint and theoretical acid dewpoint. The exact point can only be determined by empirical means, using judgment from experience. The severity of corrosion depends on the materials used.

The degree of reheat required to protect carbon steel has not been confirmed by experience, but about 30°C (50°F) is often prescribed.

Fan Protection

When dry fans are installed ahead of an FGD system, reheat does not affect fan selection. Fly ash collection must, of course, be installed ahead of the fan. Although this may raise the cost of the flue gas cleaning system, it removes the fans from the list of factors to be considered in designing for reheat.

In some installations gas is drawn through the scrubber by an induced-draft fan. To keep the fan in balance, scrubber residue must not be allowed to deposit on fan blades. A wet gas (or high dewpoint) favors such deposition. Reheat is needed to be sure that all dispersed solids in the gas stream are dry before they enter the fan intake. A reheat level of 30°C (50°F) has often been prescribed to attempt to prevent such a condition.

Wet fans and washed fans have also been used; in such cases no reheat is applied ahead of the fan. Wet fans are high in initial cost because the fan wheel is fabricated from high alloy steels, while the casing is rubber lined. The wet fan consumes less power because the wet gas is somewhat lower in volume (Ref. 4). Experience with wet fans is limited.

Plume Appearance

When the gas leaves the stack, water vapor condenses in the cooler atmosphere forming a visible plume. As the visible plume disperses, condensed vapor evaporates at a rate depending on the ambient humidity and temperature. With high external humidities early in the day, the visible plume may travel long distances before disappearing. In a desert environment at mid-day, on the other hand, the visible plume vanishes rapidly. If the appearance of such a plume is considered undesirable, it can be reduced or eliminated by reheating so that condensation does not occur.

The NSPS calling for 20 percent opacity as originally promulgated by the EPA Administrator in 1971 specifically provided that the effect of uncombined water vapor be excluded from the standard. In October 1975 this provision was stricken from the standard; however, Method 9, the procedure for visual determination, was not altered. Method 9 identifies two types of plume: attached and detached. The method specifies that the opacity be evaluated at a point following moisture plume disappearance in the case of attached plumes, and before the point of condensation in the case of detached plumes. As a practical matter this poses a difficult judgment for the observer who may not be able to discern the boundary between steam plume and that caused by fly ash. Reheat may therefore be used as a means for assuring minimum opacity readings. In the case of attached plumes, however, the provision of Method 9 favors having the maximum steam plume practicable since actual visible emissions may be masked until they are adequately dispersed.

Figure 6-1 is a psychrometric chart showing the relationship between moisture content, temperature, and plume visibility. A straight line ties the condition of the flue gas to that of the ambient atmosphere. When this line crosses the saturation curve, a visible plume will form from condensation. The longer the line on the condensation side of the curve, the more persistent the plume. Addition of heat will displace the line to the right of the saturation curve, preventing visible steam plume formation.

Stack Rain

Stack rain is a serious concern in the design of scrubber installations. Condensation in the exit duct and stack may result in mist droplets that

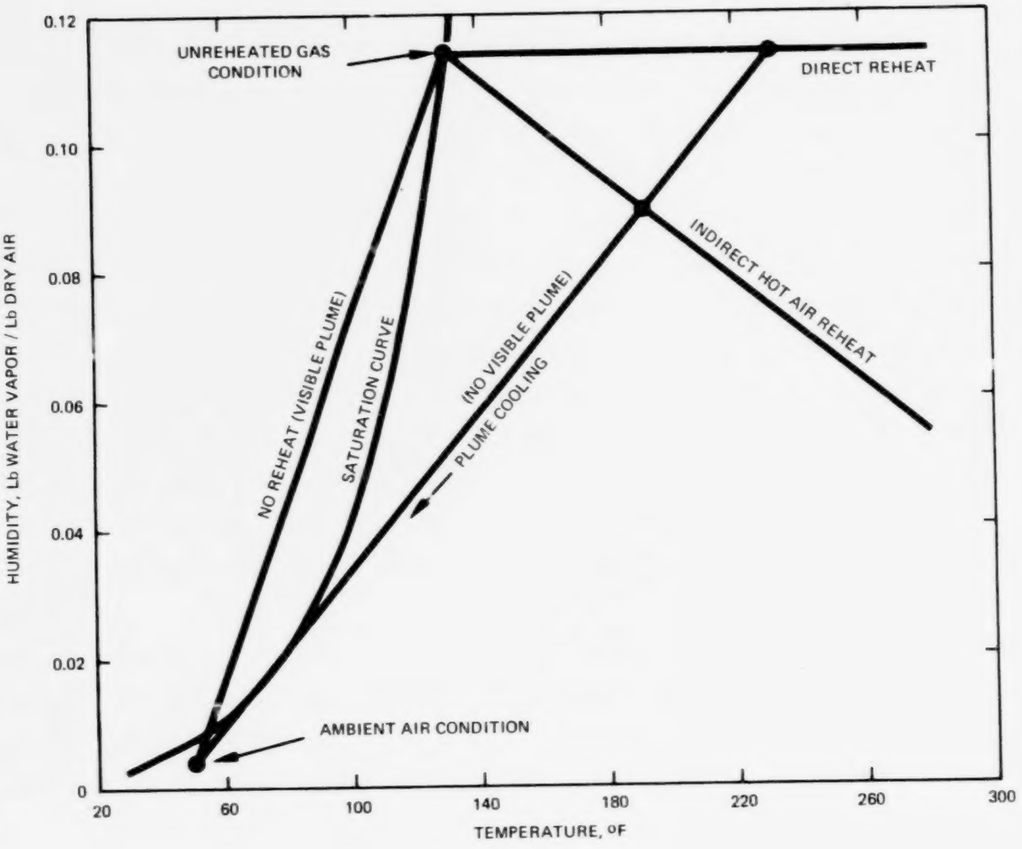

Figure 6-1. Example of Plume Behavior for Reheat Alternatives

can settle around the power station. The situation is aggravated if the mist eliminator is so inefficient that liquor blows out the stack without condensation (Ref. 5). Typical phenomena are:

- Large, muddy droplets with a pH less than 8
- Small, clean droplets, mistlike in nature, pH less than 5

The first condition is characteristic of a high velocity stack coupled with incomplete mist elimination. The second is less important and may occur only intermittently.

With the double-wall construction that typifies modern power plant stacks, the temperature loss in the stack may be as little as 3 to 6°C (5 to 10°F) but sufficient to cause condensate nucleation without agglomeration by the time the gas reaches the top of the stack. Heavy droplets are more likely to result from impingement and coalescense on the wall near the base of the stack, coupled with the drag effect of gas flow on the resulting water film at velocities higher than 12 to 15 m/sec (40 to 50 ft/sec). In the absence of reheat, wet fans and inefficient scrubber mist eliminators lead to impingement and film buildup.

Little reheat is needed to prevent the stack rain that may be caused by condensation due to stack temperature drop. Use of reheat to overcome the effect of liquid entrainment from the scrubber or from the wet fans is wasteful; use of more efficient mist eliminators would be the preferred alternative.

Stack Icing

Under freezing conditions liquid entrainment may lead to ice formation on top of the stack. The possibility of ice falling from the stack is a hazardous condition that can be alleviated by reheat.

Opacity Monitoring

The appearance of a plume can be quantitatively expressed by opacity, the plume's ability to reflect light and to interfere with light transmissions. The opacity results from the scattering of light by solid particles and condensed or entrained liquid droplets in the plume. Persistent opacity as the plume disperses is related to solids continuing after droplets have been evaporated.

Continuous optical monitoring of the flue gas opacity within the stack is a current NSPS requirement. This provides a useful tool for estimating emission and visual opacity values based on calibrated correlations with measurements taken by Methods 5 and 9. The presence of droplets because of cooling or entrainment interferes with the precision of the optical devices which do not distinguish between liquids and solids. Reheat which effects vaporization of liquids serves to overcome this drawback.

METHODS OF REHEATING

Flue gas can be reheated in many ways, and several approaches have been developed. The basic differences in reheat methods are energy sources and methods of transferring that energy to the flue gas. As shown in Figure 6-2, several different energy sources and energy transfer methods can be used.

Reheat methods currently in use include:

- Direct inline reheat — using steam or hot water heat exchangers
- Direct combustion reheat — using gas or oil in either inline burners or external combustion chambers
- Indirect hot air reheat — using steam to heat air to mix with the scrubbed gas

ENERGY SOURCES			ENERGY TRANSFER METHODS		
			HEAT EXCHANGER		MIXING
			TUBE	LJUNGSTROM	
USEFUL SYSTEM ENERGY	EXTRACTION STEAM	DIRECT	Figure 6-5	▨	▨
		INTERMEDIATE HOT AIR	▨	▨	Figure 6-6
	HOT FLUE GAS (> 300° F)	DIRECT			Figure 6-7
		INTERMEDIATE HOT AIR	▨	▨	
WASTE SYSTEM ENERGY	COLD UNSCRUBBED FLUE GAS (< 300°F)	DIRECT			Figure 6-8
		INTERMEDIATE WATER		▨	▨
EXTERNAL ENERGY	CLEAN FUEL (WITH ADDITIONAL AIR)		▨	▨	Figure 6-3
	WASTE FROM OTHER PROCESSES				Figure 6-4

▨ Not Practical

Figure 6-2. Reheat Energy Sources and Transfer Methods

- Bypass reheat — bypassing a portion of the untreated hot flue gas to mix with the scrubbed gas

Some typical approaches to reheat are shown in Figures 6-3 through 6-7.

Energy Sources

Energy may come from the power plant itself as useful or waste energy or from an external source. Figure 6-2 shows energy sources that may be used, and discussion of these sources is presented below.

Useful System Energy. Useful system energy refers to energy extracted from the power plant system itself which might otherwise be used to produce electricity. Two general streams of fluid contain such useful energy: extraction steam and hot flue gas.

Extraction steam comes from the system either before the steam turbine or at some intermediate stage of expansion before the last turbine stage. Hot flue gas refers to flue gas that retains some useful energy; for example, that between the economizer section and the air preheater section at a temperature between 260 and 400°C (500 and 750°F). Each source is available either to heat the scrubbed gas directly or to heat air which can then be mixed with the scrubbed gas.

The choice depends on convenience, optimization, and cost. In either case, extraction of energy from the system must be factored into its design. A modification is to use flue gas recirculated from a point downstream of the reheater. This would amount to bleeding off a portion of the reheated gas in Figure 6-6 and cycling it through the steam coil heat exchanger in place of the air. Since the gas is hotter than the air, there could be a savings in the amount of energy used, but costs for ducts would be higher. This method is being tested experimentally at small scale by TVA (Ref. 6).

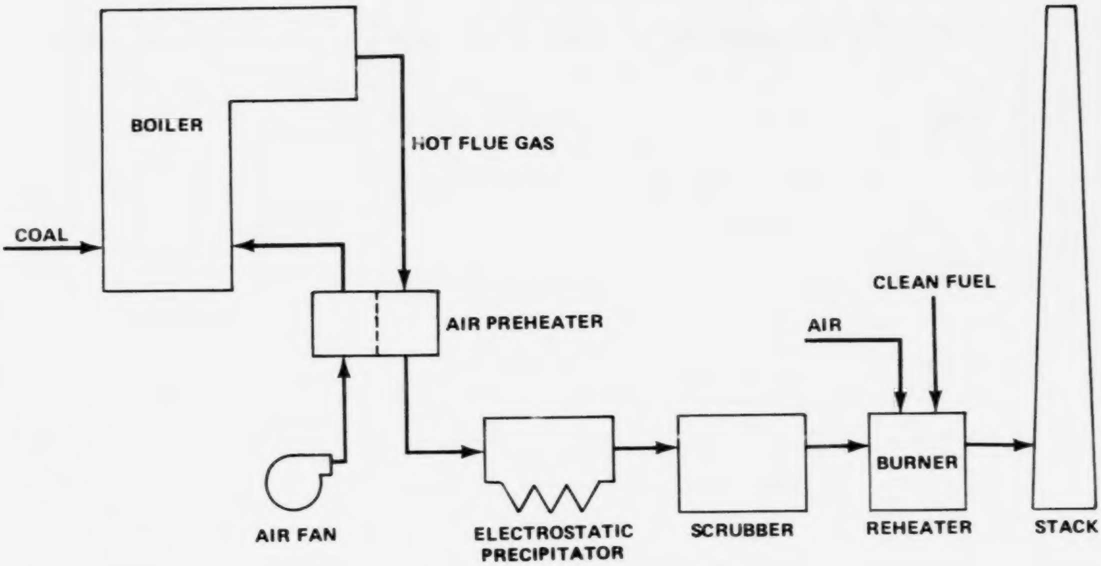
BOILER
HOT FLUE GAS
COAL
AIR PREHEATER
CLEAN FUEL
AIR
BURNER
AIR FAN
ELECTROSTATIC PRECIPITATOR
SCRUBBER
REHEATER
STACK

Figure 6-3. Direct Combustion Reheat

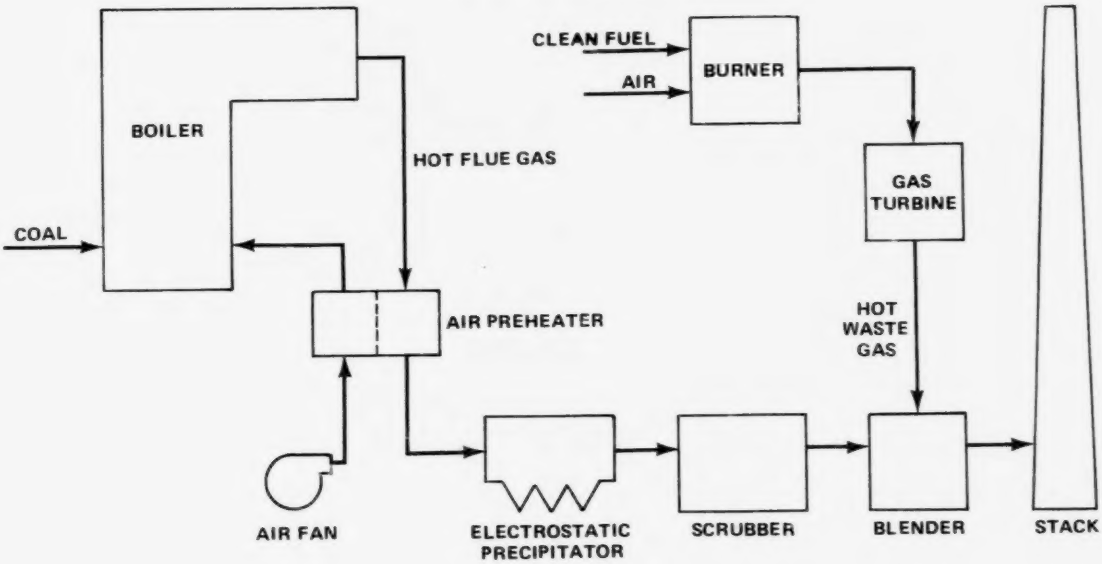
BOILER
HOT FLUE GAS
COAL
AIR PREHEATER
CLEAN FUEL
AIR
BURNER
GAS TURBINE
HOT WASTE GAS
AIR FAN
ELECTROSTATIC PRECIPITATOR
SCRUBBER
BLENDER
STACK

Figure 6-4. Reheat with External Waste Energy

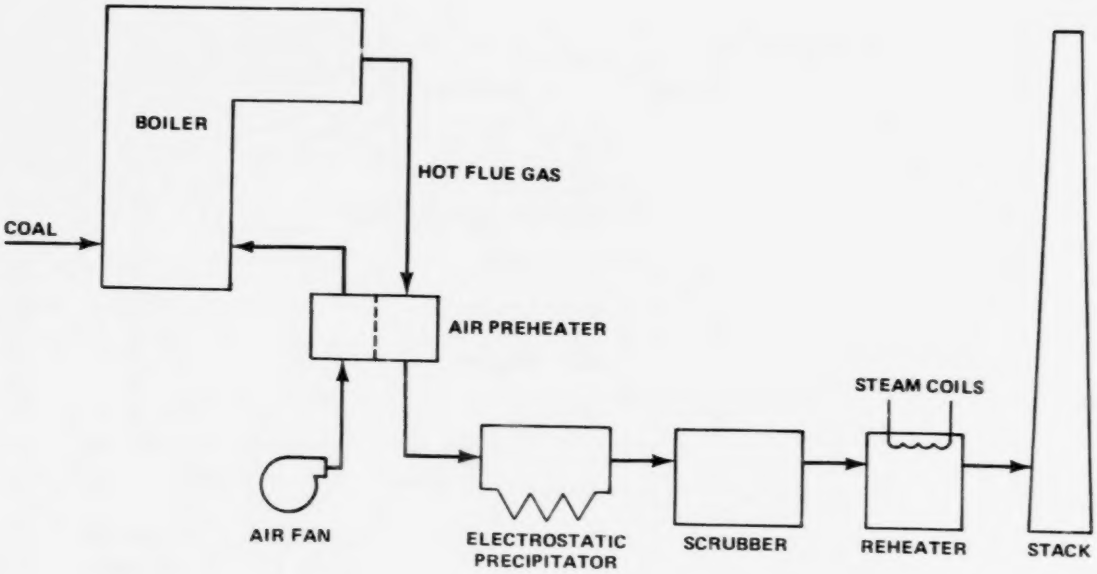
BOILER
HOT FLUE GAS
COAL
AIR PREHEATER
STEAM COILS
AIR FAN
ELECTROSTATIC PRECIPITATOR
SCRUBBER
REHEATER
STACK

Figure 6-5. Direct Inline Reheat with Steam

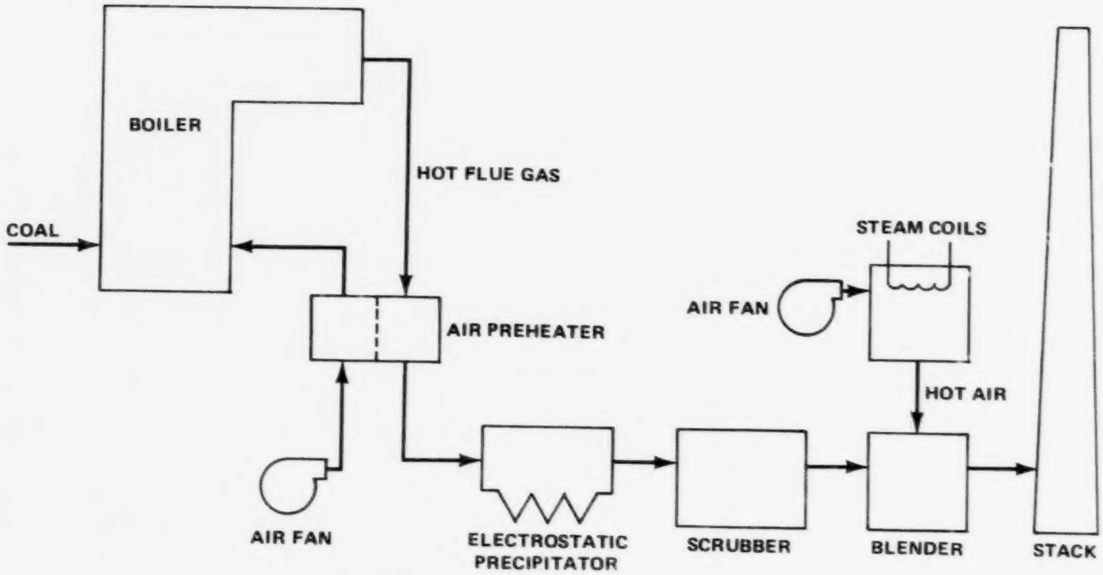
BOILER
HOT FLUE GAS
COAL
AIR PREHEATER
STEAM COILS
AIR FAN
HOT AIR
AIR FAN
ELECTROSTATIC PRECIPITATOR
SCRUBBER
BLENDER
STACK

Figure 6-6. Indirect Hot Air Reheat

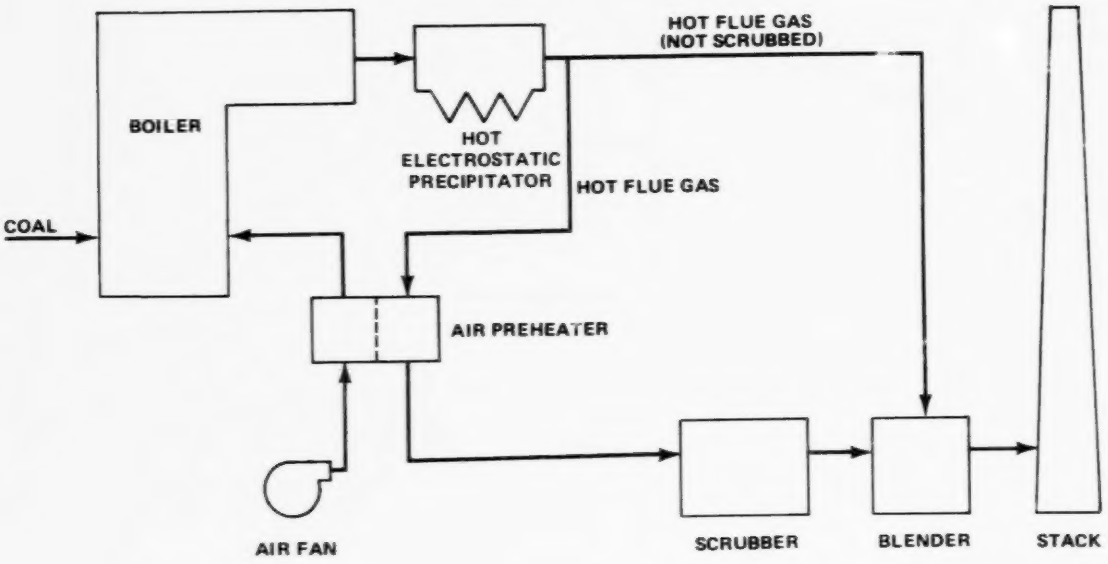
HOT FLUE GAS
(NOT SCRUBBED)
BOILER
HOT
ELECTROSTATIC
PRECIPITATOR
HOT FLUE GAS
COAL
AIR PREHEATER
AIR FAN
SCRUBBER
BLENDER
STACK

Figure 6-7. Bypass Reheat — Useful System Energy

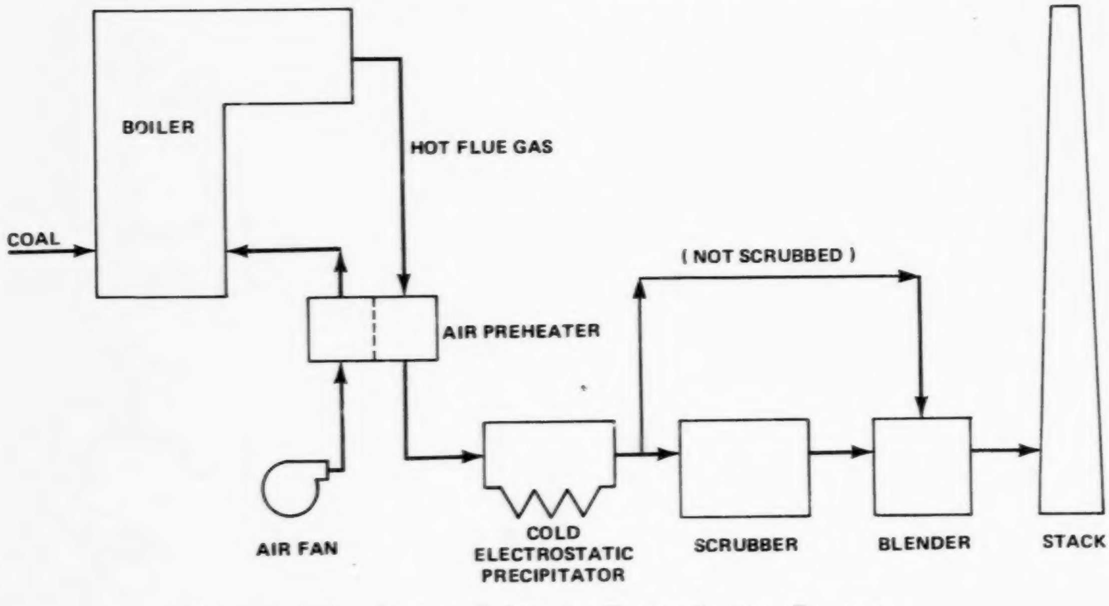
BOILER
HOT FLUE GAS
COAL
(NOT SCRUBBED)
AIR PREHEATER
AIR FAN
COLD
ELECTROSTATIC
PRECIPITATOR
SCRUBBER
BLENDER
STACK

Figure 6-8. Bypass Reheat — Waste System Energy

Waste System Energy. In a typical power plant, low-level energy is present in the warm stack gas and the condenser cooling water. Obviously, the only source of these two which is hot enough for reheat is the warm flue gas before it enters the scrubber. Some of the energy can be released either by transfer of mass directly to the scrubbed flue gas (i.e., bypass), or by using an intermediate loop with heat transfer fluid in a closed system. This fluid would convey heat from the 120 to 150°C (250 to 300°F) flue gas downstream to the scrubbed flue gas. The latter approach is being tested experimentally at small scale by TVA (Ref. 6).

External Energy. When it is not feasible to use waste energy in the fashion described above, supplemental energy sources can be used. This may be particularly applicable for retrofit installations. One method is to burn natural gas or low-sulfur oil, followed by direct injection of the combustion products into the scrubbed flue gas. The main drawback of this type of reheat in the U.S. is the limited availability and cost of the low-sulfur fuel required. Another method might use waste energy from a separate system; for example, exhaust gases from gas turbines which are generally at 315 to 370°C (600 to 700°F).

Heat Transfer Methods

The second factor to be considered, in addition to the energy source, is the method of energy transfer. Either a heat exchanger or mixing is employed.

Heat Exchanger. Two types of heat exchangers are used in power plants: the tube type and the regenerative or Ljungstrom type. Both have been used extensively, and their advantages and drawbacks are well known.

Mixing. A simpler, but not necessarily more economical, way of adding energy is to add the hot gas to the cold mixture. This technique requires care to achieve homogenous mixing.

ENERGY CONSUMPTION

Reheating consumes energy in two ways: the heat transfer required to raise the temperature of the gas and the fan energy required to force the flue gas through the reheater.

When the flue gas is forced through a heat exchanger, the energy requirement for a given temperature increment of reheat will be minimized because all of the heat transferred goes directly to heating the flue gas. On the other hand, the pressure drop through such a heat exchanger may be significant and will require some fan energy. When energy is conveyed to the flue gas by mixing it with a hotter gas, the energy requirement for a given temperature increment is generally larger than for the inline heat exchanger. This is because a larger amount of gas must be heated (flue gas plus mixing gas). In general, the pressure drop in a mixing chamber will be smaller than that in a heat exchanger.

Gas mixing methods of reheat include indirect hot air reheat (with air drawn from the outside, heated in an exchanger, and mixed with the flue gas) and direct combustion reheat (clean fuel burned in air and the combustion products mixed). They differ in their energy requirements depending on the reheat increment. In the case of indirect hot air, the volume, rather than temperature, of hot air is increased to raise the temperature of the scrubbed gas because a constant temperature steam extraction source (or hot water) is available as the primary energy source. To achieve higher temperatures with direct combustion reheat, more fuel is burned; but less energy may be required because the combustion products are at higher temperature than for the heated air method.

The energy source for reheating is an important variable. Although it may not affect the energy requirement, the electrical energy lost may well be different for different sources. For example, if waste heat is used, by definition no electricity would be lost. On the other hand, the penalty associated with steam use depends on the point in the cycle at which the energy is extracted. If steam is extracted at the highest temperature and pressure available (an unlikely choice), the cost of that heat will be very high. If steam is extracted at some intermediate point, the cost is lower. In determining the cost of such heat, its value in terms of available electrical energy must be considered. The boiler and turbine must be designed with the reheat steam penalty in mind.

Bypass reheat has the advantage of low capital investment, negligible operating cost, and simple and reliable operation. However, the maximum degree of reheat obtainable is limited by the overall SO_2 removal requirement versus the SO_2 removal capability of the FGD system.

The relative theoretical energy requirements for various degrees of reheat have been calculated by Battelle (Ref. 6). Results of these calculations are shown in Figure 6-9. The results are for a typical coal-fired boiler. In actual practice, the energy consumption may be slightly higher than that indicated because of mist eliminator inefficiency (mist carryover and latent heat penalty).

It should be noted that for the same degree of reheat, indirect hot air reheat has the highest energy requirement. However, for the same amount of energy consumption, indirect hot air reheat may have equal or better benefits because of its dilution effect on the plume.

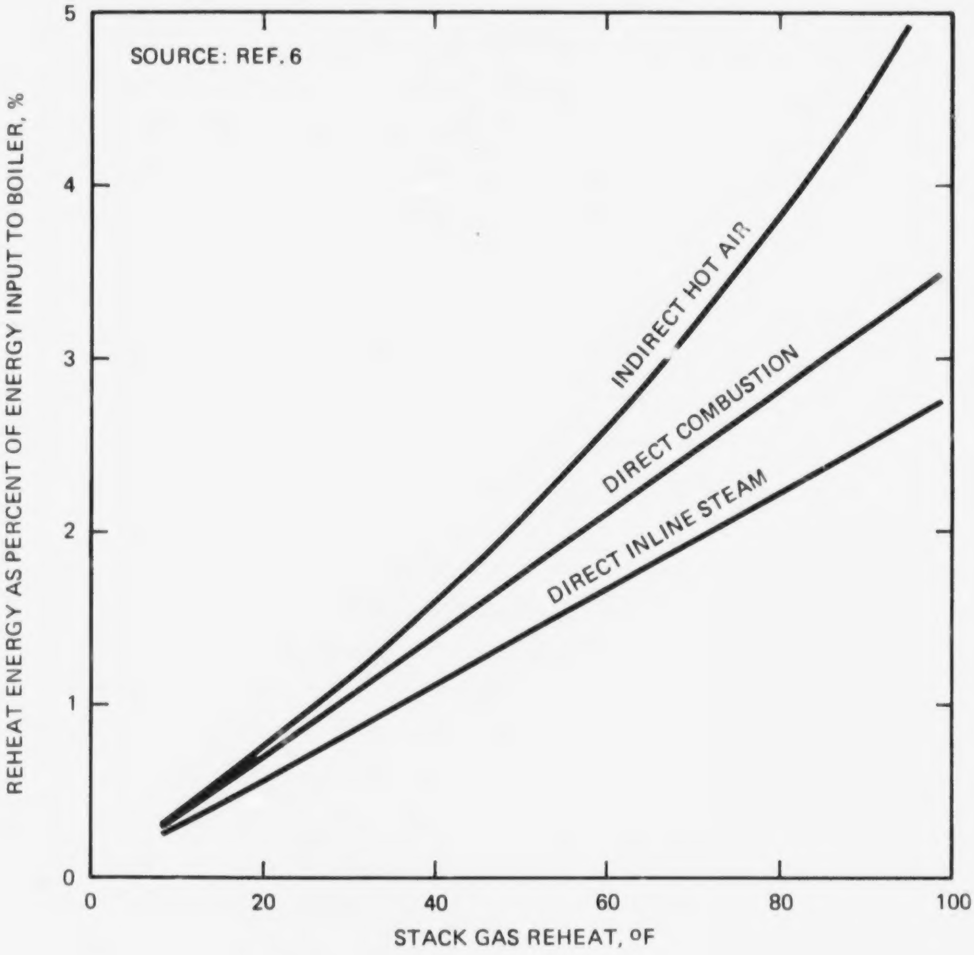

Figure 6-9. Theoretical Reheat Energy Requirements for Typical Coal-Fired Boiler

PROBLEMS EXPERIENCED

This section examines some of the reheater problems experienced in operational flue gas desulfurization systems. Tables 6-1 and 6-2 indicate the type of reheat in each generating station considered and Tables 6-3 through 6-6 highlight the problems within each system. It is of interest to note that the following power plants operate without reheat:

- Cane Run No. 4
  Louisville Gas and Electric Company
- Green River 1 and 2
  Kentucky Utilities
- Conesville No. 5
  Columbus and Southern Ohio Electric Company
- Duck Creek No. 1A
  Central Illinois Light Company
- Southwest 1
  Springfield City Utilities

Bypass reheat is employed by Martin Lake 1, Texas Utilities Company, and additional FGD systems are under construction with provision for bypass reheat. Operating experience to date is limited, but the selection of duct lining material for the hot-cold gas interface area is a critical design factor. The material must be capable of withstanding thermal shock, the high temperature of the bypassed gas stream, and the corrosive effects of the scrubbed gas.

Table 6-1 classifies the heat exchanger type of reheat into three subdivisions:

- Direct steam
- Intermediate hot air
- Direct hot water

The direct steam reheater consists of a heat exchanger installed in the duct containing the flue gas. Condensing steam inside the heat exchanger heats the flue gas to the desired reheat temperature.

Table 6-1

REHEAT USING HEAT EXCHANGERS

Power Plant	Type of Reheat	Heating Source	Tube Style	Materials of Construction	Tube Diameter, Inches	Degree of Reheat, ΔT°F
La Cygne,(a) Kansas City Power and Light	Useful Energy, Direct	Steam @ 140 psig & 690°F	Bare	316LSS(b)	0.625	70
Will County Commonwealth Edison	Useful Energy, Direct	Steam @ 350 psig & 485°F	Bare	Bottom Section, 316LSS(c) Top Section, Carbon Steel(c)	0.625	70
Cholla, Arizona Public Service	Useful Energy, Direct	Steam @ 250 psig & 405°F	Bare	316LSS	1.0	60
Mohave Vertical Module, Southern California Edison	Useful Energy, Direct	Steam @ 600 psig & 650°F	Bare	"E-Brite" 26 Chromium 1-Molybdenum	1.0	80
Mohave Horizontal Module, Southern California Edison	Useful Energy, Intermediate Hot Air	Steam @ 600 psig & 650°F	Finned	Carbon Steel	1.0	70
Colstrip, Montana Power	Useful Energy, Direct	Steam @ 150 psig 360°F	Plate	Bottom Section Inconel 625 Top Section, Hastelloy G	(d)	30
Valmont-5, Public Service Company of Colorado	Useful Energy, Direct	Steam @ 450 psig 900°F	Finned	Carbon Steel	1.5	70
Hawthorn, Kansas City Power and Light	Useful Energy, Direct	Hot Water @ 325°F	Finned	Carbon Steel	1.0	50
Lawrence, Kansas Power and Light	Useful Energy, Direct	Hot Water @ 250°F	Finned	Carbon Steel	1.0	30
Sherburne County, Northern States Power	Useful Energy, Direct	Hot Water @ 350°F	Finned	Carbon Steel	0.5	40
Four Corners, Arizona Public Service	Useful Energy, Intermediate Hot Air	Steam @ 600 psig & 650°F	Finned	Carbon Steel	1.0	50
Reid Gardner, Nevada Power	Useful Energy, Intermediate Hot Air	Steam @ 450 psig & 760°F	Finned	Carbon Steel	0.625	50

(a) Heated air is also injected below the heat exchanger.
(b) Original material of construction was 304SS.
(c) Original material of construction was 304 SS and Corten.
(d) Does not apply to plate type heat exchangers.

Table 6-2

REHEAT USING COMBUSTION GAS

Power Plant	Location of the Combustion Chamber	Type of Fuel	Degree of Reheat $\Delta T^{o}F$
Eddystone 1A, Philadelphia Electric	External	Oil	100
Phillips, Duquesne Power	Internal	Oil	30
Elrama, Duquesne Power	Internal	Oil	30
Bruce Mansfield, Pennsylvania Power	Internal	Oil	NA
St. Clair 6, Detroit Edison	External	Oil	100
Paddy's Run 6, Louisville Gas and Electric	Internal	Gas	40
D. H. Mitchell* Northern Indiana Public Service	Internal	Gas	50

*This reheat system has not yet been put in service.

Intermediate hot air for reheat is attained by passing outside air through condensing steam in a heat exchanger and mixing with flue gas.

Direct hot water reheat is similar to the inline steam approach, except that lower temperature differences are involved. The heat transfer fluid is pressurized hot water.

Flue gas and intermediate air heat exchangers are normally tubular in shape. Extended surface fins have been used to enhance heat transfer. The following materials of construction have been used (see Table 6-1): carbon steel, 304 stainless steel, 316 (low carbon) stainless steel, Corten, Inconel 625, and Hastelloy G. The relative effectiveness of these materials for reheater service is discussed in the following section. Typically, the tubes are 1 inch or less in diameter. Plate coils have also been used.

The degree of reheat for the installations listed in Tables 6-1 and 6-2 varies from 17 to 56°C (30 to 100°F). The amount of reheat required is site specific, and there is no known correlation between generating stations.

The problems associated with each method of reheat are distinct and will be discussed as related to the following categories:

- Direct steam and hot water reheat (Tables 6-3 and 6-4)
- Intermediate heated air reheat (Table 6-5)
- Gas and oil fired reheat (Table 6-6)

Direct reheaters have failed because of plugging, corrosion, and vibration. Plugging occurs from entrainment. Once the scrubber liquor deposits on the reheater, the dissolved and suspended solids bake onto it. This deposit continues to grow, blocking the gas flow, increasing the pressure drop, and helping to induce corrosion from the localized high temperature and concentrated salts. The effectiveness of the heat transfer surface is progressively reduced as the deposit builds up.

Table 6-3

DIRECT STEAM REHEAT OF SCRUBBED FLUE GASES

Power Plant	Reheater Problem	Solution Attempted	Solution Follow-Up
La Cygne, Kansas City Power and Light (7,8)	The 304SS reheater failed because of chloride stress corrosion and acid condensate corrosion.	Replaced the 304SS reheater tubes with 316LSS tubes.	Not available
		Hot air (supplied from the boiler's air heater) was injected below the reheater to prevent acid condensation.	
		Increased the reheater flue gas temperature from 147°F to 175°F. The utility expects that 316LSS will withstand corrosion at 175°F.	
	The reheater plugged with scrubber slurry passed through the mist eliminator.	Modified and corrected the operating procedures relating to the mist eliminator and reheater (details not available).	Intermittent heavy over spray and continuous underspray have kept the mist-eliminator relatively clean.
		Modified the steam soot blower.	The soot blower modification aided in keeping the reheater clean.
		One module is cleaned each night (there are seven modules).	Not available
Will County, Commonwealth Edison (9,10,11)	Module A Reheater: Reheater tubes failed because of chloride corrosion (1972-1974). The reheater also plugged with solids passed through the mist eliminator.	No corrective action taken at the time of tube failure. Module A reheater was put back in service by cannibalizing the Module B reheater.	The corrective actions relieved but did not completely eliminate corrosion.
		In addition, the mist eliminator wash method was improved by constant underspray and an intermittent overspray.	Wyoming low sulfur coal was burned during 1974; there were very few problems. Once the station started burning high sulfur coal in mid-1975, the mist eliminator and reheater problems reappeared and the module was shut down for the balance of 1975.
		The mist eliminator washwater was changed from pond reclaim water to service water.	
		Steam was injected during scrubber outages to slow down the tube deterioration.	During mid-1975 the reheater and mist eliminator were replaced.
		A second stage demister was installed above the original demister to prevent washwater carryover to the reheater.	
	Module B Reheater: Reheater tubes failed because of vibration during high gas flow conditions (1972).	The reheater was rebraced and a baffle installed to reduce the vibrations.	Not available
	The reheater failed because of chloride corrosion (1973). Note: Module B was shut down April, 1973 and not restarted until May, 1975.	The Module B reheater was rebuilt based on the experience gained on Module A. 316SS and carbon steel tubes replaced the 304SS and Corten tubes.	Carbon steel reheater tubes failed.

Table 6-3 (Cont'd)

Power Plant	Reheater Problem	Solution Attempted	Solution Follow-Up
	Reheater tubes failed because of vibration (<1,000 hours operation). The tube supports had worked loose permitting the tubes to vibrate.	The tube supports were redesigned and the carbon steel tube bundles repaired.	Not available.
	The carbon steel tube support frames corroded.	No information available.	
	Reheater plugging occurred in the same manner as Module A.	The same solutions were followed for both Modules A & B.	No reported reheater plugging even when the mist-eliminator scaled and plugged while burning high sulfur coal in mid-1975.
Cholla, Arizona Public Service (12,13,14)	Module A Reheater: Harmonic vibrations with deflections as much as 0.040 inches occurred in the reheaters. The vibration was attributed to improper flue gas distribution.	Baffles were installed.	The baffles reduced but did not eliminate the vibration, but it was no longer considered a major problem.
	Little information is available on Module A reheater corrosion. Most corrosion problems occurred in the Module B reheater.		
	Module B Reheater: Harmonic vibrations also occurred in the Module B reheater.	Baffles were installed.	The baffles reduced but did not eliminate vibration.
	Both chlorides and sulfurous acid corroded the reheater. Sulfurous acid formed on the duct walls and dripped onto the reheater tubes. Chloride attack occurred when entrained water containing chlorides impinged on the tubes.	To prevent sulfurous attack the ducts above the reheater were insulated. In addition a trough was built around the duct to catch any run off before it reached the reheater. No solution postulated for chloride corrosion.	It is the intent that new reheat bundles would be of Inconel 625 rather than the original 316LSS.
	The reheater housing also corroded.		
Mohave Vertical Module, Southern California Edison (15)	There was no problem with the vertical module reheater other than a few seams failing on the seamless rated tubing. No reheater plugging occurred during the testing.	None	Upon completion of the testing, the tubes were inspected. Pits were detected in the "E-Brite" 26-Cr 1 Mo tubes.

Table 6-3 (Cont'd)

Power Plant	Reheater Problem	Solution Attempted	Solution Follow-Up
Colstrip 1 and 2, Montana Power (16)	No problems have been reported with the inconel 625 and Hastelloy G plate type reheaters. Some loose scale formed on the reheaters but did not cause any operating problems.		
Valmont 5, Public Service Company of Colorado (14,16)	No reported reheater plugging or corrosion during short-term SO_2 removal tests. The scrubbers are normally used for particulate removal only.		

Table 6-4

DIRECT HOT WATER REHEAT

Power Plant	Reheater Problem	Solution Attempted	Solution Follow-Up
Hawthorn, Kansas City Power and Light (10,14,17)	Hawthorn No. 3 Reheater No major reheater problem has been reported.		
	Hawthorn No. 4 Reheater No reheater corrosion problem reported. Reheater plugging has been a problem particularly in the B side of the scrubber.	A section of the reheater was removed to facilitate cleaning and maintenance.	It is normal practice to shut down the scrubber every three days for cleaning of the mist eliminator. At that time the reheater may be cleaned.
Lawrence, Kansas Power and Light (6,10,14,18,19)	Lawrence No. 4 Reheater This reheater plugged frequently during the early life of this system.	To alleviate the plugging problem the following changes were made: a. Soot blowers were installed beneath the reheater. b. The mist eliminator was redesigned. c. Vanes were installed under the marble bed to improve gas distribution.	The soot blower pressure flattened the copper fins. The copper finned tubes were replaced with carbon steel finned tubes. Mist eliminator plugging continued to be a recurring problem.
	Carbon steel reheater completely failed (1976).	This scrubber will be completely razed and replaced by a venturi rod scrubber followed by a spray tower (to be completed in 1977).	
	Lawrence No. 5 Reheater No problems reported on this reheater. The corrections made on No. 4 were incorporated into the Lawrence No. 5 scrubber system.		
Sherburne, Northern States Utilities (14)	There has been multiple failure of the carbon steel finned tubes.	No information available	
	Reheater plugging is a common event.	The scrubbers are shut down on a three-day cycle for cleaning.	

Table 6-5

STEAM HEATED AIR REHEAT

Power Plant	Reheater Problem	Solution Attempted	Solution Follow-Up
Mohave Horizontal Module, Southern California Edison (20)	There were no reported problems with this reheat system.		The scrubber system was moved to the Four-Corners Generating Station for additional testing.
Four-Corners, Arizona Public Service (21)	Horizontal Scrubber Reheater: The carbon steel tubes developed pinholes. Metallurgical examination revealed that the pinholes developed from inside the tube as a result of stagnant condensate and air that leaked into the reheat system. It is not known when the pinholes developed but the system had been idle for a year prior to testing. The pinholes probably developed during this idle period.	The pinholes were brazed so that the system testing could continue. A nitrogen blanket system was installed in the reheater loop to exclude air from the inside of the tube.	The horizontal scrubber has been shut down indefinitely. There is no feedback as to the nitrogen blanket performance.
Reid Gardner, Nevada Power (14,22)	Reheater leaks have been reported but their severity has not been indicated.	No information available	

Table 6-6

GAS AND OIL FIRED REHEAT

Power Plant	Reheater Problem	Solution Attempted	Solution Follow-Up
Eddystone 1A, Philadelphia Electric (14,23,24)	The combustion chamber double brick refractory lining failed. The failure occurred because of torch placement and problems with the ultra violet flame scanners. Also a portion of the duct was insulated which limited heat dissipation and caused subsequent failure of the refractory and steel sheet.	No information available.	
Phillips, Duquesne Light (6,14,25,26,27)	Problems occurred with the oil pumps, burners and temperature control. Corrosion has been reported in the combustion chamber.	No information available. (No priority has been placed on operating the reheater because of the oil shortage.)	A new acid proof stack liner has been installed. This reheat system has hardly been operated. The scrubber functions with a wet stack.
Elrama, Duquesne Light (14,27)	Problems similar to the Phillips Power Station's.	No information available.	
Bruce Mansfield No. 1, Pennsylvania Power (28)	Combustion problems have occurred because of the flue gas mixing with the combustion air (i.e., problems in maintaining a flame in the combustion chamber. Reheater vibration is also a problem.	These problems are to be resolved by the equipment supplier.	No information available.
St. Clair 6, Detroit Edison (6,14)	The only problem reported is the failure of a thermal controller resulting in liner damage. In addition, the firebricks in the flue gas mixing zone fell off because of vibration.	No information available.	
Paddy's Run 6, Louisville Gas and Electric (14)	No reported problems		
D. H. Mitchell, Northern Indiana Public Service (30,31)	Reheat system has not been put in service.		

Corrosion originating in this manner has led to serious failures. The scrubber liquor usually contains dissolved chlorides that concentrate within the deposits and attack the metal. Stainless steel, particularly type 304, is vulnerable to stress corrosion in the presence of chloride. Gaseous SO_2 also slowly attacks carbon steel and Corten above the acid dewpoint and rapidly below. Sulfurous acid from this source will condense on cold reheater tubes (when operated without steam or hot water) and on cold duct surfaces during startup or low load periods. If the reheater is in a vertical duct, acid condensate formed on the duct will run down the duct onto the reheater tubes.

Reheaters have failed from vibration fatigue. If the natural frequency of the reheater matches that of the fan, reheater tubes may wear against their support baffles and fail.

The major problem with hot air injection systems, to date, has been condensate corrosion. Typically, the failure has been with carbon steel, a particularly vulnerable material.

With oil-fired systems the problems appear in the course of attempting to maintain the flame within the main flue gas stream. This saves space, but typically does not work well. Refractory failures have occurred in the combustion chamber from flame impingement, attack from condensation during downtime, or other reasons. Brick linings have also failed from vibration. Most refractories are weak in tensile strength, and both rapid heating and vibration tend to cause failure.

Burning oil in the presence of flue gas causes complications in lighting off the burner, flame-outs during operation, and formation of soot. Burning natural gas in the flue gas stream is reported to have been done successfully at the Louisville Gas and Electric Paddy's Run Station.

DESIGN, OPERATION AND MAINTENANCE OF REHEATERS

The previous section outlined typical problems that have occurred in operating installations with reheaters. Avoiding these problems is best done by making use of operating experience.

No particular method of reheat can be endorsed for general use. A well designed, operated, and maintained reheat system should be able to function with a minimum of problems. Selection of a system is increasingly a matter of sound design.

Design, operation, and maintenance of reheaters fall into the three categories already identified:

- Direct steam and hot water
- Intermediate hot air
- Gas and oil combustion gas

Direct Steam and Hot Water Reheaters

Direct steam and hot water reheaters have experienced three major problems: plugging, corrosion, and vibration.

Reheater Plugging. As described in the previous section, reheater plugging is the result of scrubber liquor containing suspended and dissolved solids passing through the mist eliminator and impinging on the reheater tubes or plates. The purpose of the mist eliminator is to prevent the scrubber liquor from leaving the scrubber. If the mist eliminator allows scrubber liquor to pass, then it requires improved design, operation, or maintenance. Proper mist eliminator design is essential to prevent plugging. The following design parameters are critical:

- Velocity through the mist eliminator
- Material of construction

- Blade spacing (if any)
- Number of stages
- Distance between the upper stage of the scrubber and the mist eliminator, or location within the scrubber

The Colstrip and Mohave stations are reported to have been relatively successful with direct steam reheater operation by correct integration of the above factors into the mist eliminator design.

Most direct reheaters incorporate a steam or air soot blower to clean any plugs that form. The quantity of steam or air used and the frequency of cleaning (most stations clean at least once every 8 hours) vary from station to station. Soot blowers are a standard feature for maintaining clean reheat surfaces; operating cycles are best derived from experience.

The best method to alleviate plugging is to minimize entrainment.

Corrosion. Direct steam and hot water reheaters have suffered repeated construction material failure because of corrosion. Operating experience suggests that certain materials should not be used, such as carbon steel, 304SS, 316SS, and Corten. More exotic materials such as Inconel 625 and Hastelloy G have been used successfully at Colstrip. The use of these materials would approximately double the cost of a 316LSS reheater.

So far, corrosion and pitting have been attributed to periodic acid conditions and to chlorides. Effective design of the mist eliminator will minimize these.

Condensation can be avoided by keeping the reheat bundles hot and by insulating ductwork downstream of the reheater. Provisions may be desirable in future designs to keep the reheater warm during prolonged outages.

Vibration. Design against vibration is a structural matter depending on the span of the tubing and its section modulus. A frequency analysis is indispensable for proper design.

Operation and Maintenance. A well designed scrubber system has ample lighting and access doors near the mist eliminator and reheater for inspection and maintenance. Warning instruments are of great value.

The soot blower cleaning cycle must be designed with enough latitude to meet actual service requirements.

Intermediate Hot Air Reheaters

Intermediate hot air reheaters have had fewer problems than inline types. The reheaters are outside the flue gas duct so that slurry never reaches the reheater surface. In one case corrosion of tubes has been experienced (see Table 6-5).

Flue gas leakage into dead-ended cold reheaters can be avoided by installing shutoff dampers between the in-service duct and the idle reheater and by purging the idle zone with air or gas. In addition, low pressure steam can be bled into the tubes to keep them hot. With such precautions, even carbon steel may demonstrate acceptable service life for some installations.

The temperature of the hot air before mixing is higher than 200°C (400°F) and can destroy the usual coatings used to protect ducts and scrubber walls. Hot air must be prevented from entering the system without cold gas flow. This requires suitable interlocks.

The same precaution should be exercised as with inline steam coils in preventing tube vibration in the air heater.

Because the reheat equipment is outside the flue gas duct, hot air injection and maintenance can be somewhat simpler than for the direct system.

Oil- and Gas-Fired Reheaters

The major problems in oil-fired reheaters have been those of arrangement — installing the combustion chamber too close to or inside the flue gas duct. Oil-fired reheat systems should be designed with an external combustion chamber so that complete combustion takes place before the hot gas mixes with the flue gas, and the provision should be made for off-line light-off; e.g., short stack.

A flame temperature of 1,600°C (3,000°F) can warp metals, destroy coatings, and crack refractories. Combustion chambers must be amply sized, and interlocks must be provided so that the reheater cannot be started unless there is flue gas in the duct and the scrubber liquor is circulating. Careful specification of refractory and subcomponents is essential.

There have been no significant problems reported with the gas-fired reheat system at Paddy's Run.

ALTERNATIVES TO REHEAT

A number of alternatives to reheating are available. Whether or not they are feasible must be considered on a case by case basis. A major factor in their applicability is whether a scrubber is being installed as a retrofit or is being designed for a new plant.

Tall Stacks

To the extent that excessive ground concentration of pollutants is a problem (due to reduced plume buoyancy from scrubbing), one alternative to reheating is to build a taller stack. A taller stack could be more economical than reheating even though it involves a high capital cost. There is, by comparison, no energy cost. However, under certain circumstances the stack height required for a particular location might not be feasible. Meteorological modeling is a useful tool for determining the validity of such an alternative, but most dispersion models have not been developed for wet plumes.

Corrosion Protection

To limit corrosion, one may either select materials that are inherently resistant to corrosion or use coatings to cover materials subject to corrosion.

Corrosion resistant materials include some costly alloys, such as the high-chromium stainless steels. Nonmetallic materials resistant to wet corrosion include fiberglass or fiberglass reinforced polyesters (FRP) for use as linings for stacks and critical sections of ductwork. However, there is an upper temperature limit on such materials in the range of 100 to 150°C (200 to 300°F), well below that of stainless steel.

A number of noncorrosive coatings and linings include materials that can be sprayed on, painted on, or prefabricated and bonded in place. Polyesters, silicate-based inorganics, epoxies, and other plastics are available. Rubber linings have long been used. Each has advantages and disadvantages based on the material performance and cost for the specific application.

A key problem is that of quality control of application of the lining. The surface preparation and bonding must be timed for periods of warmth and low humidity. Once applied, bonding adequacy is very difficult to determine. If bond failure occurs and corrosion is discovered, repairs can be costly. Further progress in the areas of lining bond quality measurement and corrosion detection is desirable.

Upstream Fan

If the purpose of reheat is to protect a downstream fan, an obvious alternative is to place a fan upstream of the scrubber. This is only feasible with an upstream collector or precipitator to remove erosive particulate matter.

Mitigation of Plume Appearance

There is very little, other than reheat, that is effective in mitigating the vapor plume. However, plume appearance and, in particular, the length of a visible plume, are strong functions of atmospheric conditions.

Figures 6-10 and 6-11 show the calculated behavior of plumes for a Southwestern United States 2,000 MW coal-fired power plant with sulfur dioxide scrubbers (Ref. 32). These figures show the frequency of plumes of various lengths. For both figures, Curve A shows the frequency distribution of plumes of a given length in the absence of reheat. It may be noted, for example, that only 11 percent of the time are plumes longer than 1 kilometer. Thus, if visibility is considered a problem only for plumes longer than a few kilometers, reheat may be unnecessary except for a small fraction of the time.

These same figures also show the effect of reheat to a given temperature on the frequency distribution of visible plumes. Figure 6-10 refers to direct reheat, meaning in this case any form of reheat that does not add mass to the flue gas, while Figure 6-11 refers to indirect reheat where it is assumed that air is first heated, then mixed with the flue gas. These curves show that a substantial increase in temperature is required to affect the frequency distribution significantly. For example, to reduce the frequency of plumes as long as 1 kilometer from 11 percent to 2 percent, it is necessary to apply almost 95°C (170°F) of reheat. It can also be seen from these figures that for a given amount of reheat, the indirect method is more effective in reducing or mitigating visible water vapor plumes. However, it must be kept in mind that the indirect application requires a higher energy input for the same amount of temperature difference. It turns out that the same effect on visibility requires approximately the same energy input.

The frequency distribution curves do indicate, however, that if the purpose of reheat is a cosmetic one (reduction of the frequency of visible plumes), a logical technique would be to use variable reheat. That is, reheat could be limited to those periods of atmospheric conditions during which a plume of objectionable length would otherwise occur.

Mist Elimination

As indicated previously, absence of reheat may be accompanied by rain or stack ice formation. Reheat, however, is costly and not necessarily the best answer to such problems. A preferred alternative would be to use more efficient mist eliminators to remove droplets of large size. Heat transfer calculations can show that reheat consumes very large amounts of energy before reducing significantly the size of large droplets.

One form of liquid emission that would not be affected by better mist elimination results from condensate forming on the inside of the stack and brought out of the stack by the rising flue gas. Such an effect can be reduced or eliminated by the combination of insulation on the stack to reduce condensation likelihood, plus the addition of liquid catchers on the upper portion of the stack.

It can also be shown that the formation of large droplets in the atmosphere by agglomeration and growth is highly unlikely in the absence of large droplet emission from the stack. Such formation should not contribute significantly either to stack icing or rain.

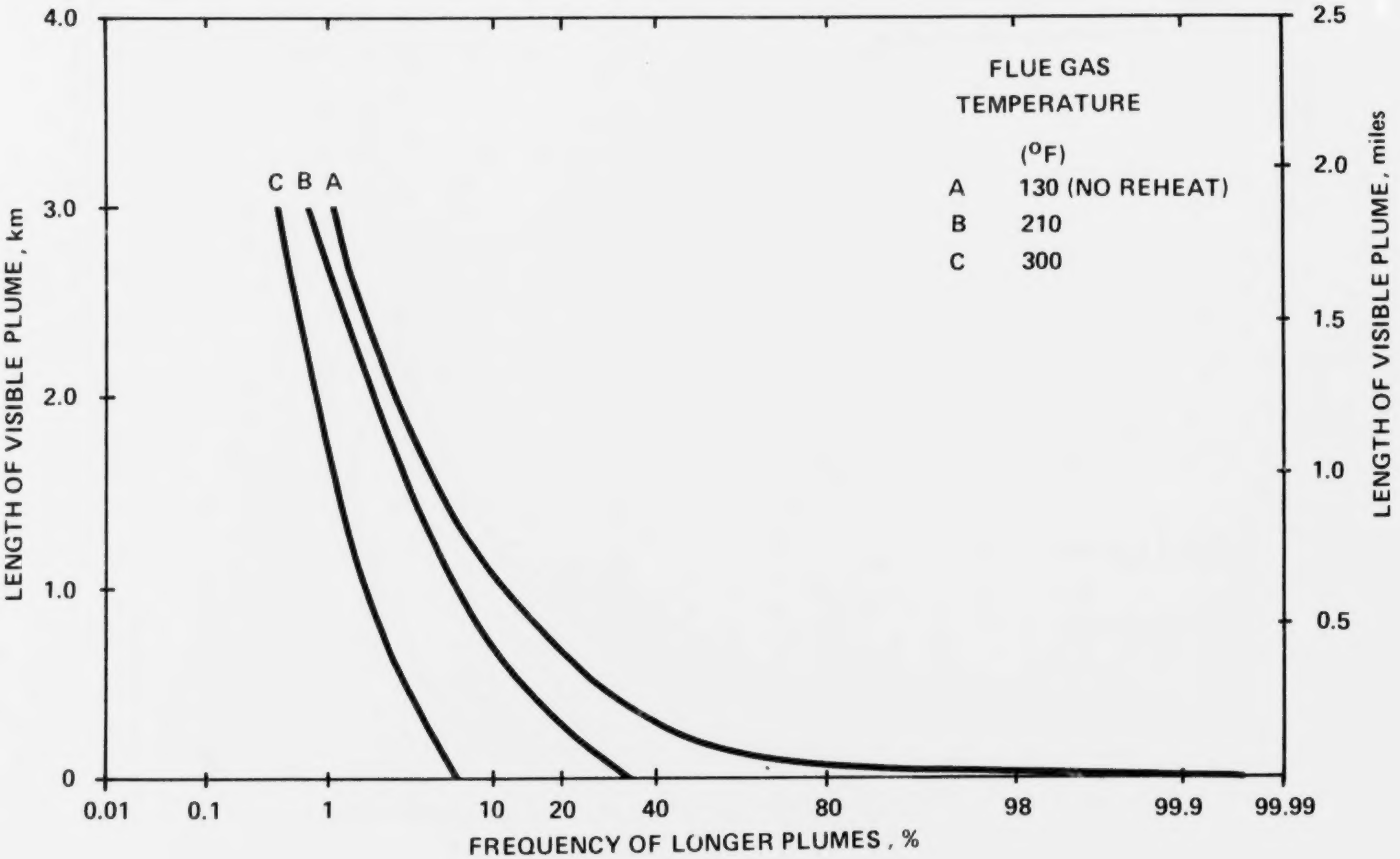

Figure 6-10. Frequency of Occurrence of Visible Plumes with Direct Reheat

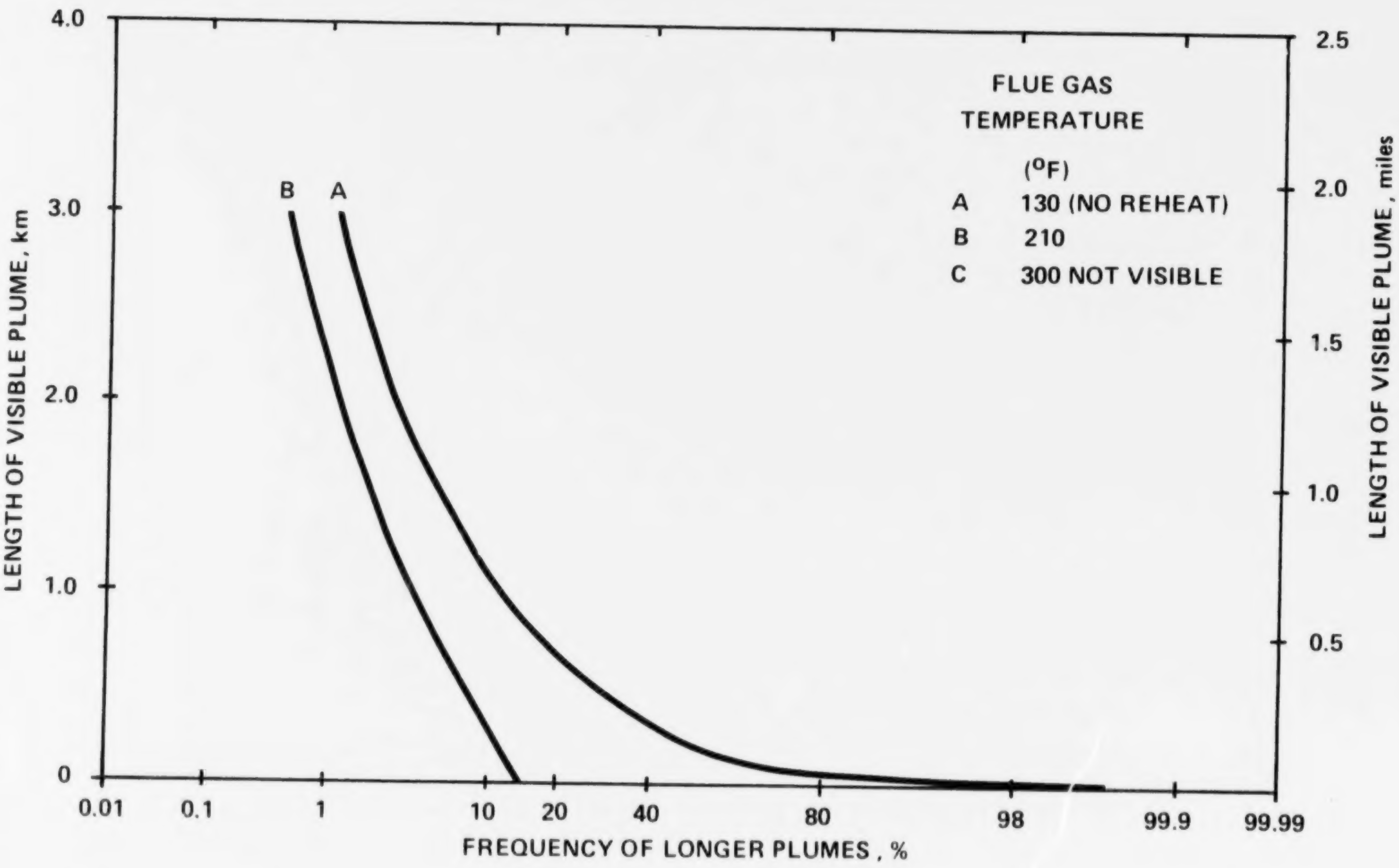

Figure 6-11. Frequency of Occurrence of Visible Plumes with Indirect Reheat

Opacity Measurement Technology

It has been suggested that one value of reheat lies in the fact that it simplifies the measurement of particle emissions without interference from liquid droplets or aerosols. While that is true, it would probably be possible to measure particle concentration even in the presence of condensed moisture. Techniques have been devised that employ an optical sensing device in a reheated bleed stream, for example. If all other arguments for reheating were eliminated, using it solely to measure opacity would constitute the most expensive opacity measurement of all.

REHEAT SUMMARY

This section has shown that while reheat of flue gas may have significant benefits, it is clearly expensive. In some cases, particularly with low sulfur, reheat can be more expensive than the alkali absorbing costs themselves. In addition, it has been shown that most of the benefits can be realized by alternative methods. Reheating should not be considered as a necessity, but as one of a number of approaches for consideration in optimizing sulfur dioxide absorption systems.

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Appendix A

DATA FROM EPA ALKALI SCRUBBING TEST FACILITY

These data were obtained at the EPA Alkali Scrubbing Test Facility located near Paducah, Kentucky. The test facility is integrated into the flue gas ductwork of a 150-MW coal-fired boiler at the TVA Shawnee Power Station. Two parallel wet scrubber systems are in operation: a venturi/spray tower system and a Turbulent Contact Absorber (TCA) system. Each system can treat approximately 10 MW equivalent (33,000 Nm^3/hr, or 30,000 acfm @ 300^oF) of flue gas containing 1,500 to 4,500 ppm of SO_2. Figures A-1 and A-2 (drawn with major dimensions to scale) show the two scrubber systems. The cross-sectional area of the spray tower is 6 m^2 (50 ft^2) in both the scrubbing section and in the mist eliminator. The cross-sectional area of the TCA scrubber is 3.0 m^2 (32 ft^2) in the scrubbing section and 4.6 m^2 (49 ft^2) in the mist eliminator section.

Figures A-3 through A-27 show the results of short-term factorial tests 6 to 8 hours in length and long-term tests about one week in length.* The short-term tests were aimed at determining SO_2 removal efficiencies without regard necessarily for possible long-term scaling conditions that might be associated with the process conditions chosen. Therefore, these particular data are not to be interpreted as evidence of long-term reliable operation. However, during other portions of the Shawnee program adequate demonstrations have been made that these FGD systems can be operated for extended periods without problems with respect to scaling.

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For the spray tower and TCA data, the SO_2 removal curves in the figures have been produced by mathematical models which are presented in Appendix B.

VENTURI SCRUBBER AND SPRAY TOWER

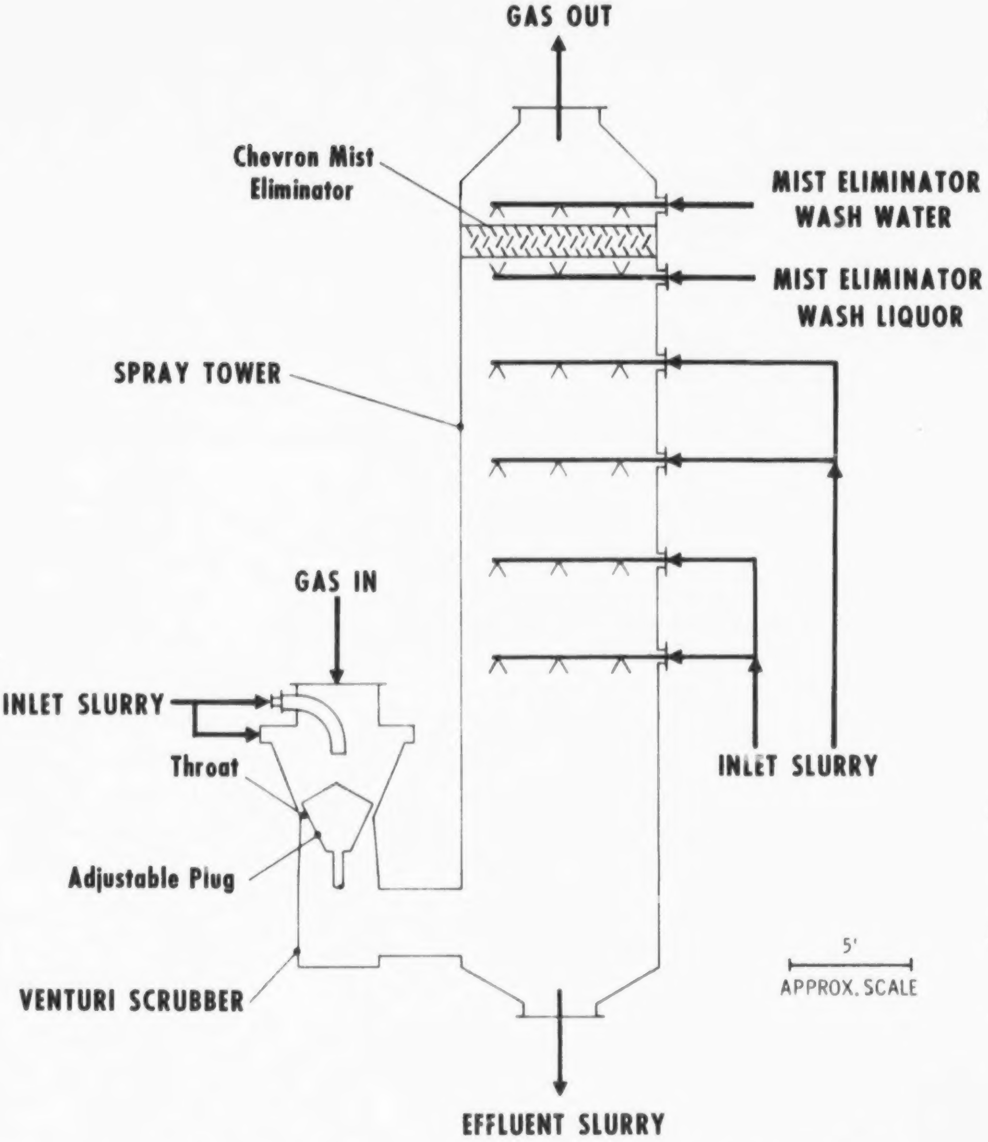
GAS OUT
Chevron Mist Eliminator
MIST ELIMINATOR WASH WATER
MIST ELIMINATOR WASH LIQUOR
SPRAY TOWER
GAS IN
INLET SLURRY
Throat
Adjustable Plug
INLET SLURRY
5'
APPROX. SCALE
VENTURI SCRUBBER
EFFLUENT SLURRY

Figure A-1. Schematic of Venturi Scrubber and Spray Tower

TCA SCRUBBER

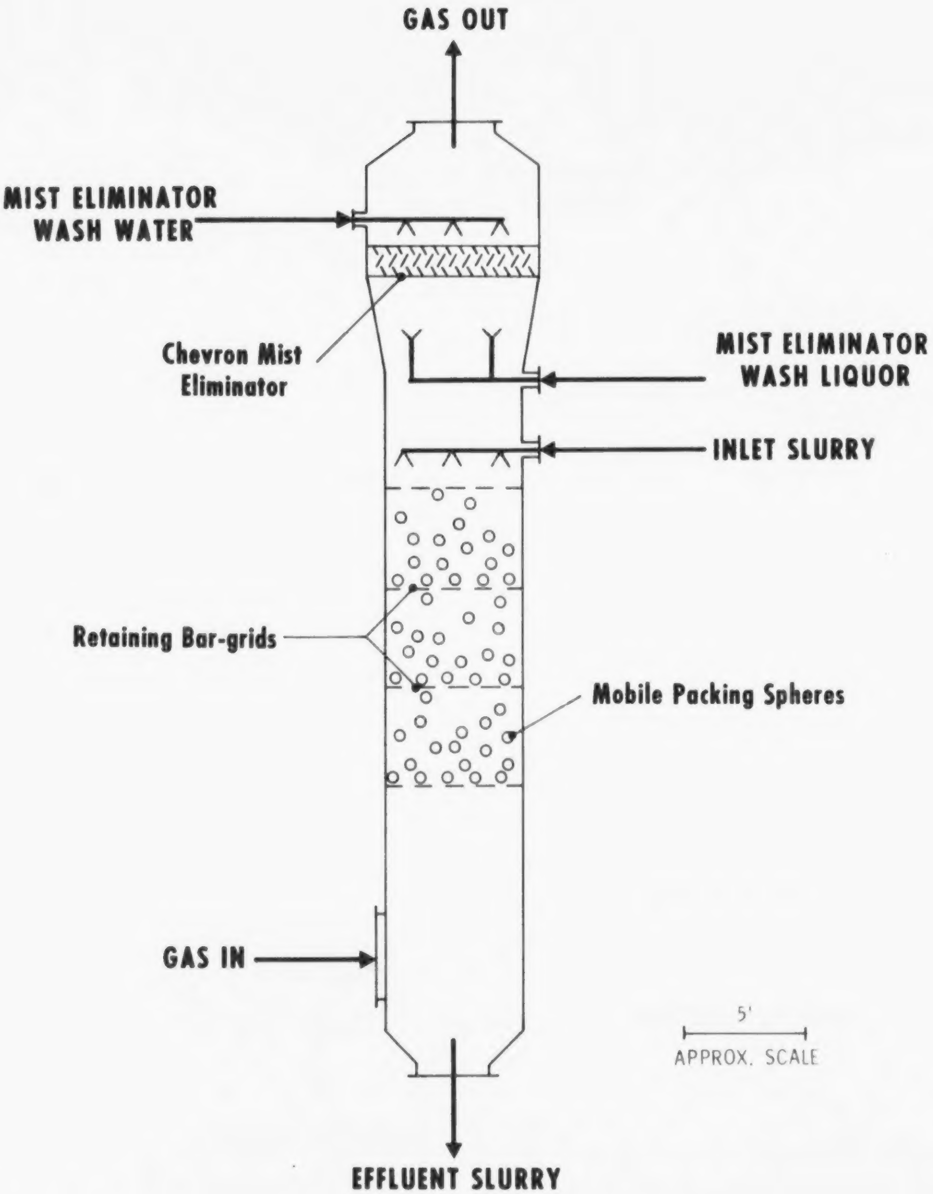
GAS OUT
MIST ELIMINATOR WASH WATER
Chevron Mist Eliminator
MIST ELIMINATOR WASH LIQUOR
INLET SLURRY
Retaining Bar-grids
Mobile Packing Spheres
GAS IN
5'
APPROX. SCALE
EFFLUENT SLURRY

Figure A-2. Schematic of Three-Bed TCA

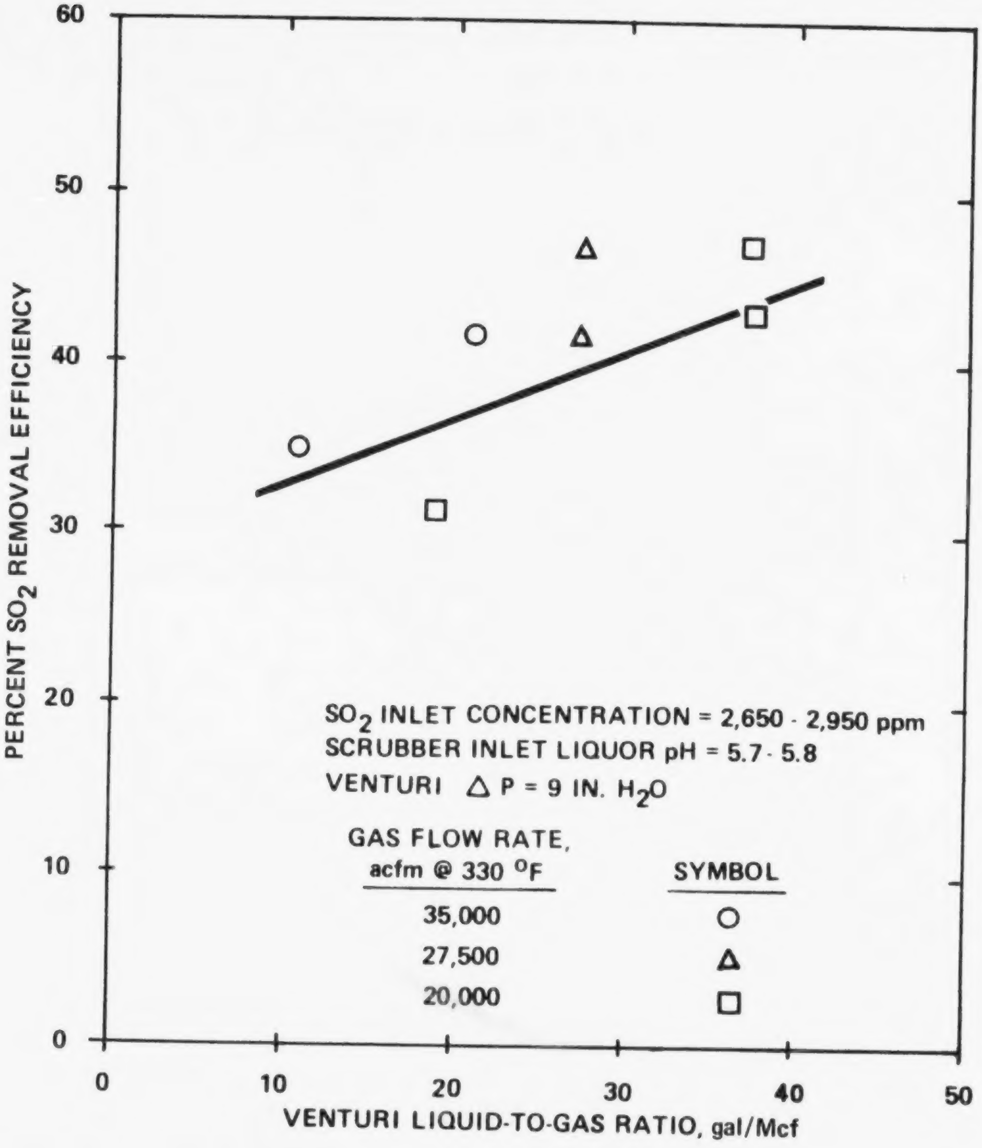

Figure A-3. Effect of Liquid-to-Gas Ratio and Gas Flow Rate on SO_2 Removal — Venturi Scrubber with Limestone

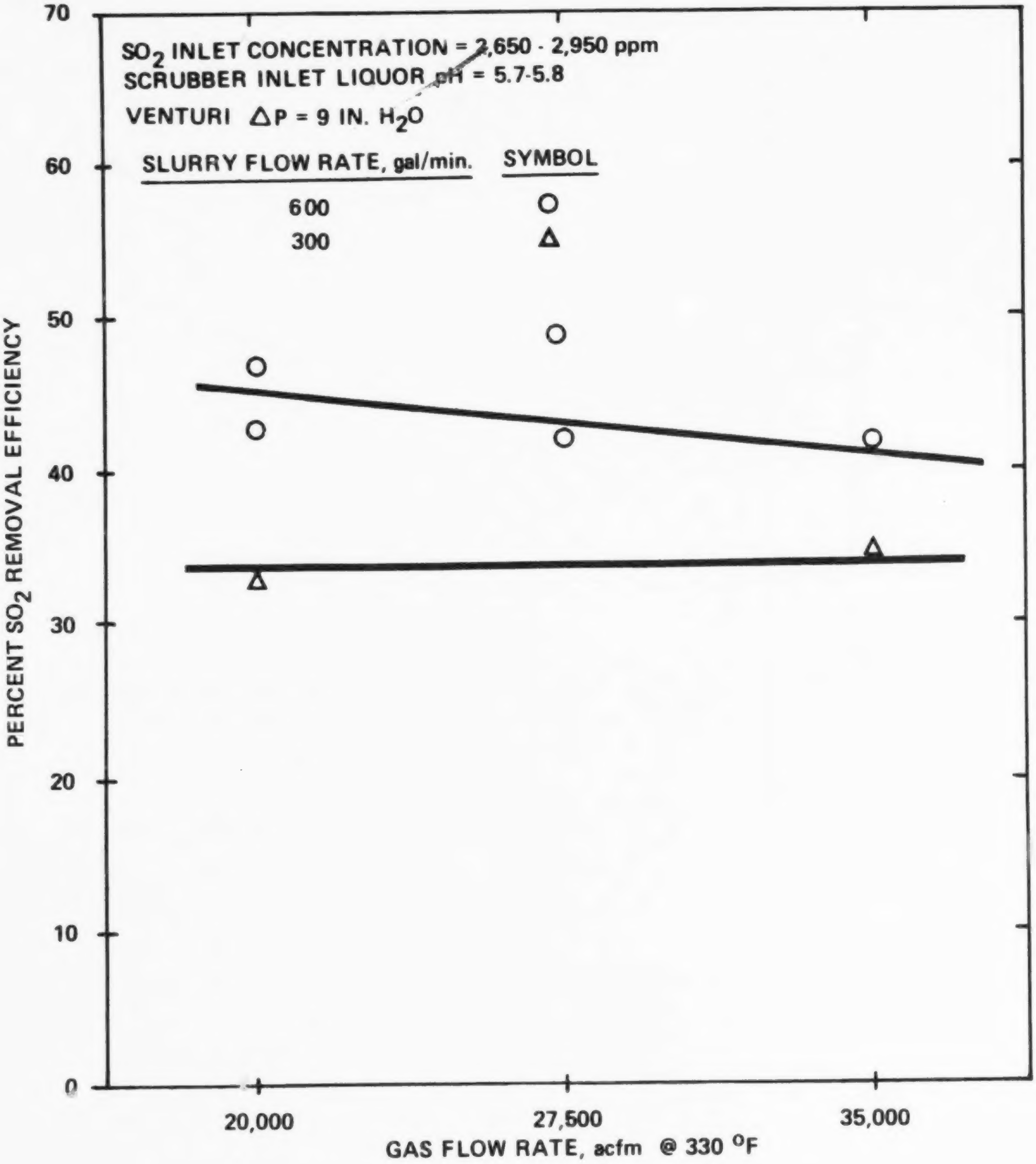

Figure A-4. Effect of Slurry and Gas Flow Rates on SO_2 Removal — Venturi Scrubber with Limestone

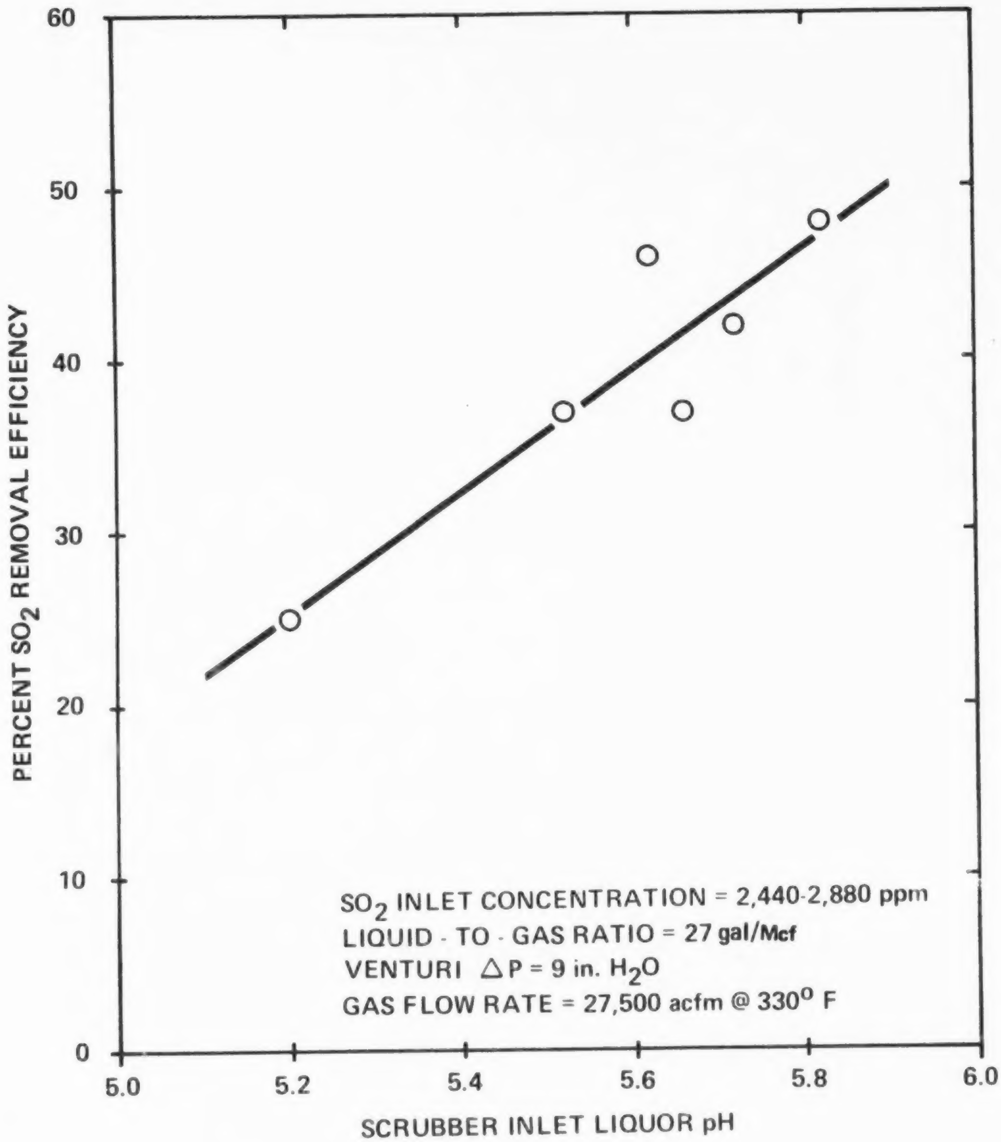

Figure A-5. Effect of Scrubber Inlet Liquor pH on SO_2 Removal — Venturi Scrubber with Limestone

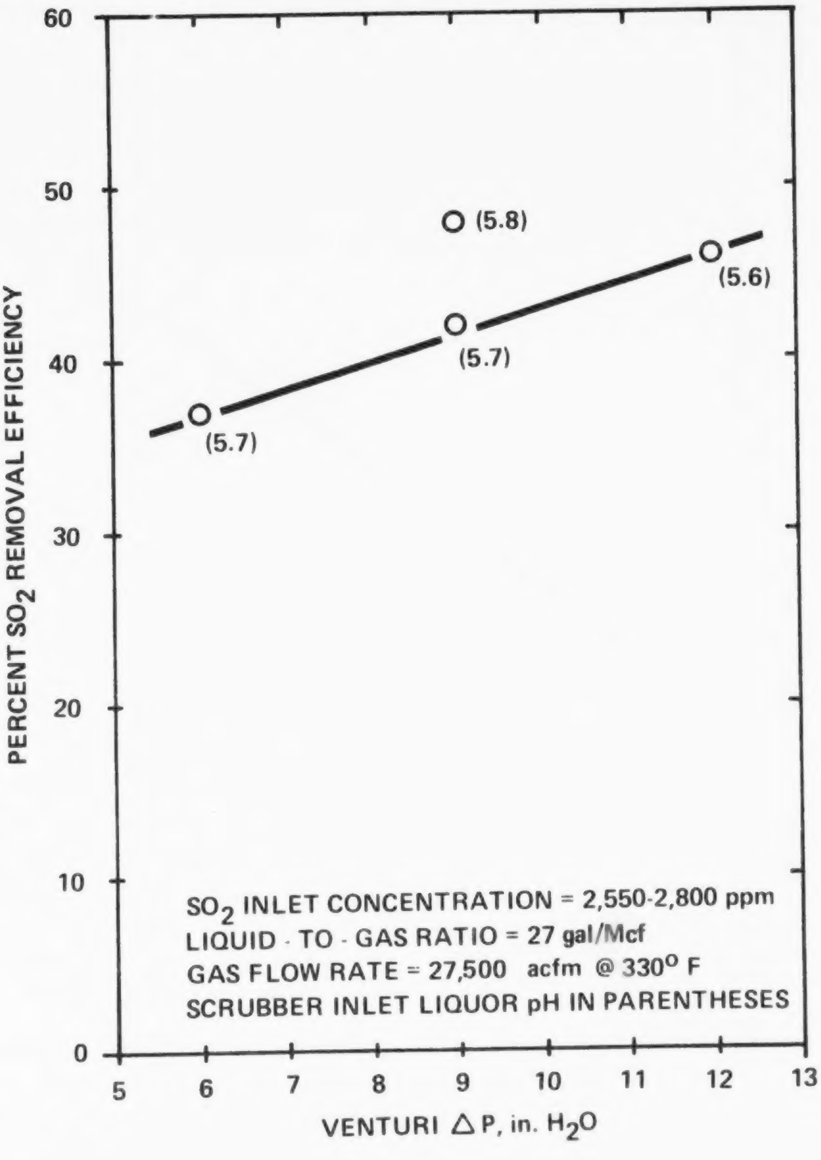

Figure A-6. Effect of Venturi ΔP on SO_2 Removal — Venturi Scrubber with Limestone

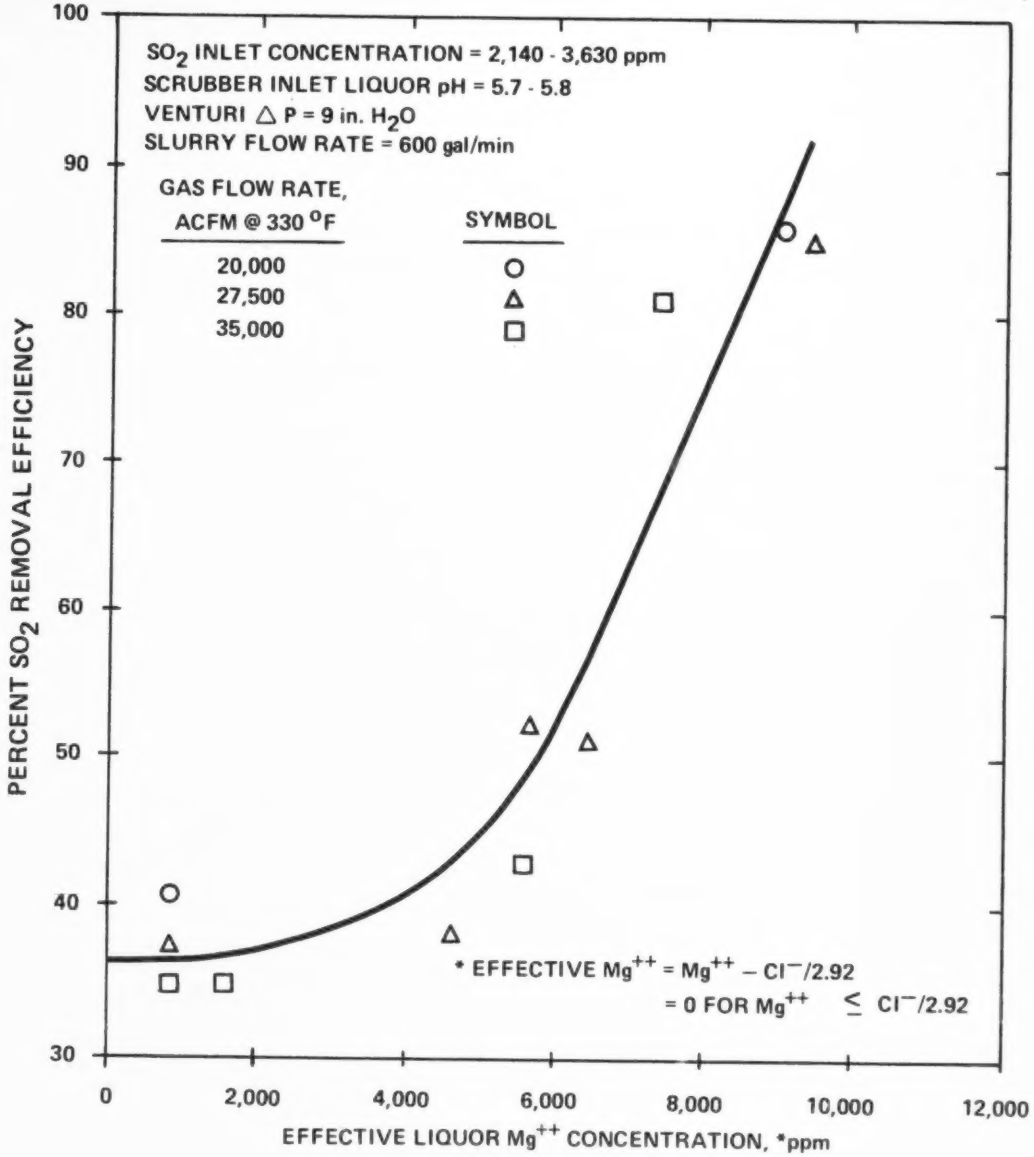

Figure A-7. Effect of Liquor Magnesium-Ion Concentration and Gas Flow Rate on SO_2 Removal — Venturi Scrubber with Limestone

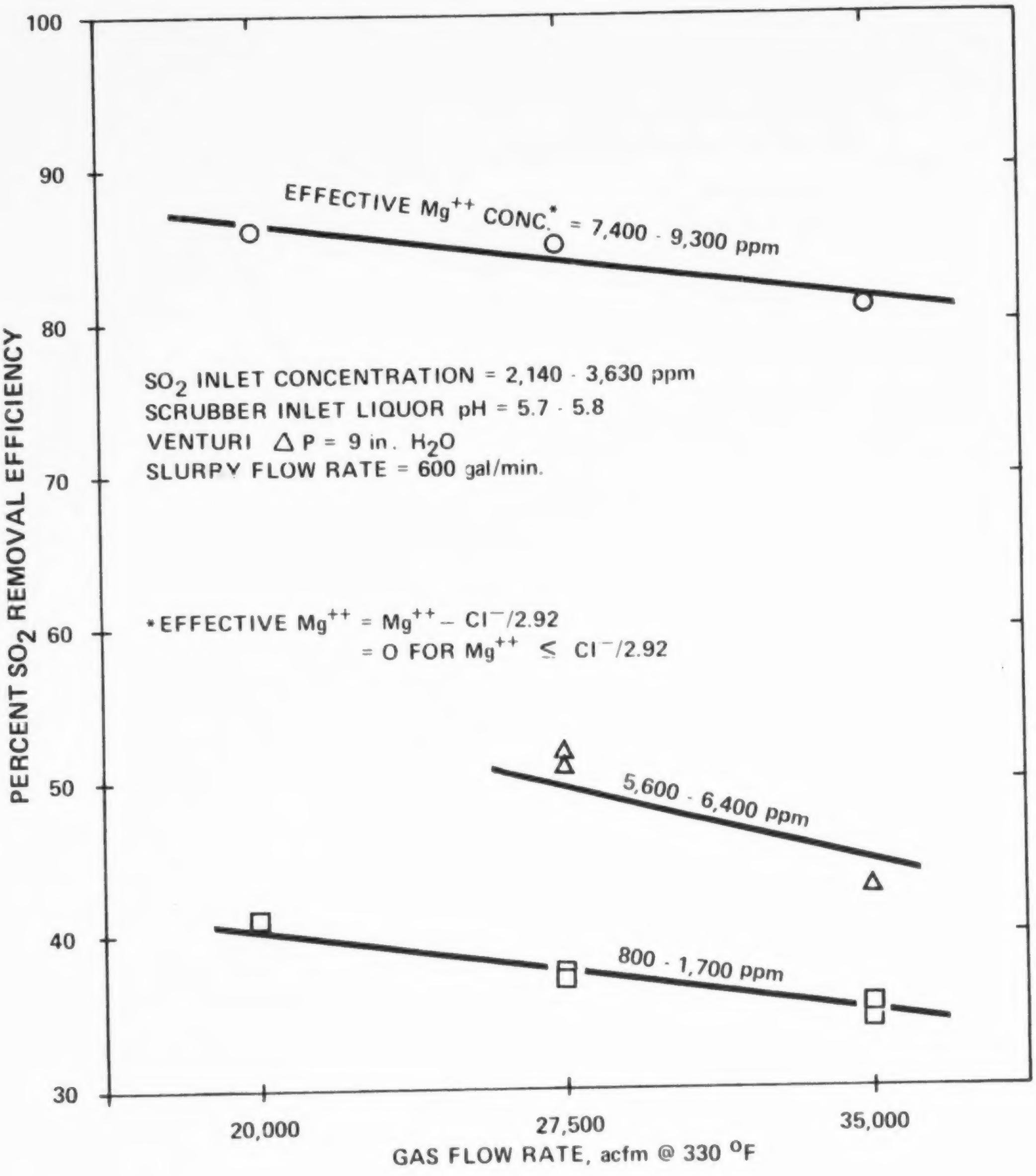

Figure A-8. Effect of Gas Flow Rate and Liquor Magnesium-Ion Concentration on SO_2 Removal — Venturi Scrubber with Limestone

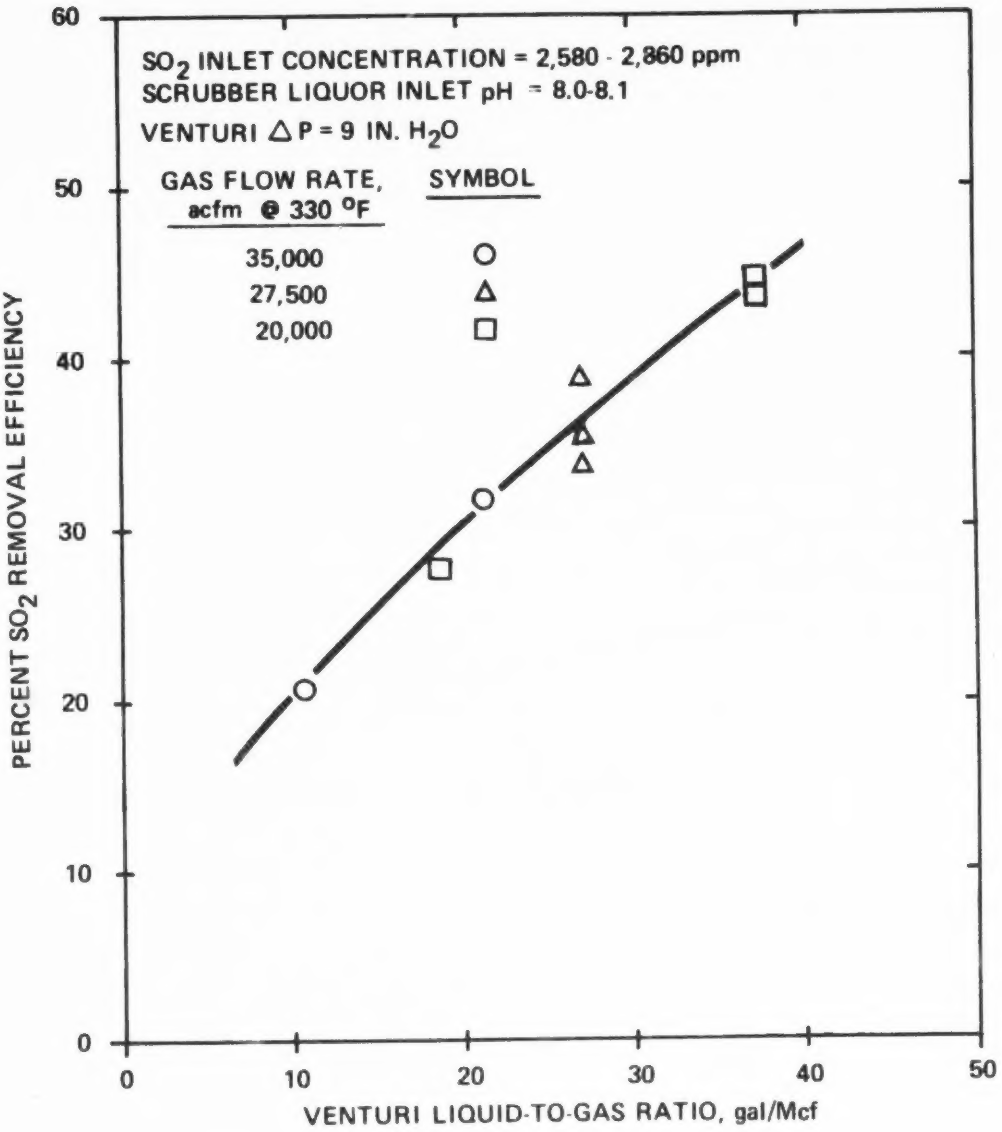

Figure A-9. Effect of Liquid-to-Gas Ratio and Gas Flow Rate on SO_2 Removal — Venturi Scrubber with Lime

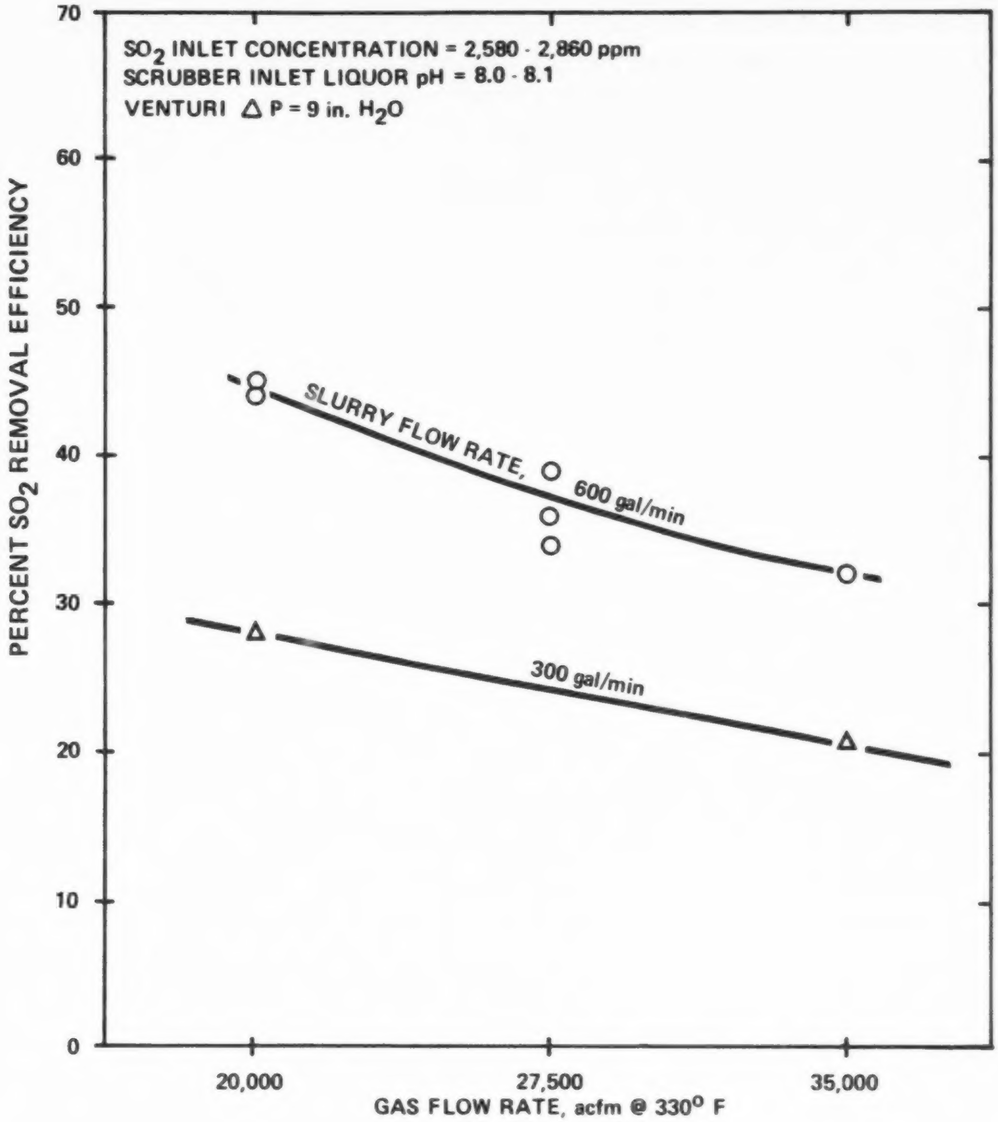

Figure A-10. Effect of Slurry and Gas Flow Rates on SO_2 Removal — Venturi Scrubber with Lime

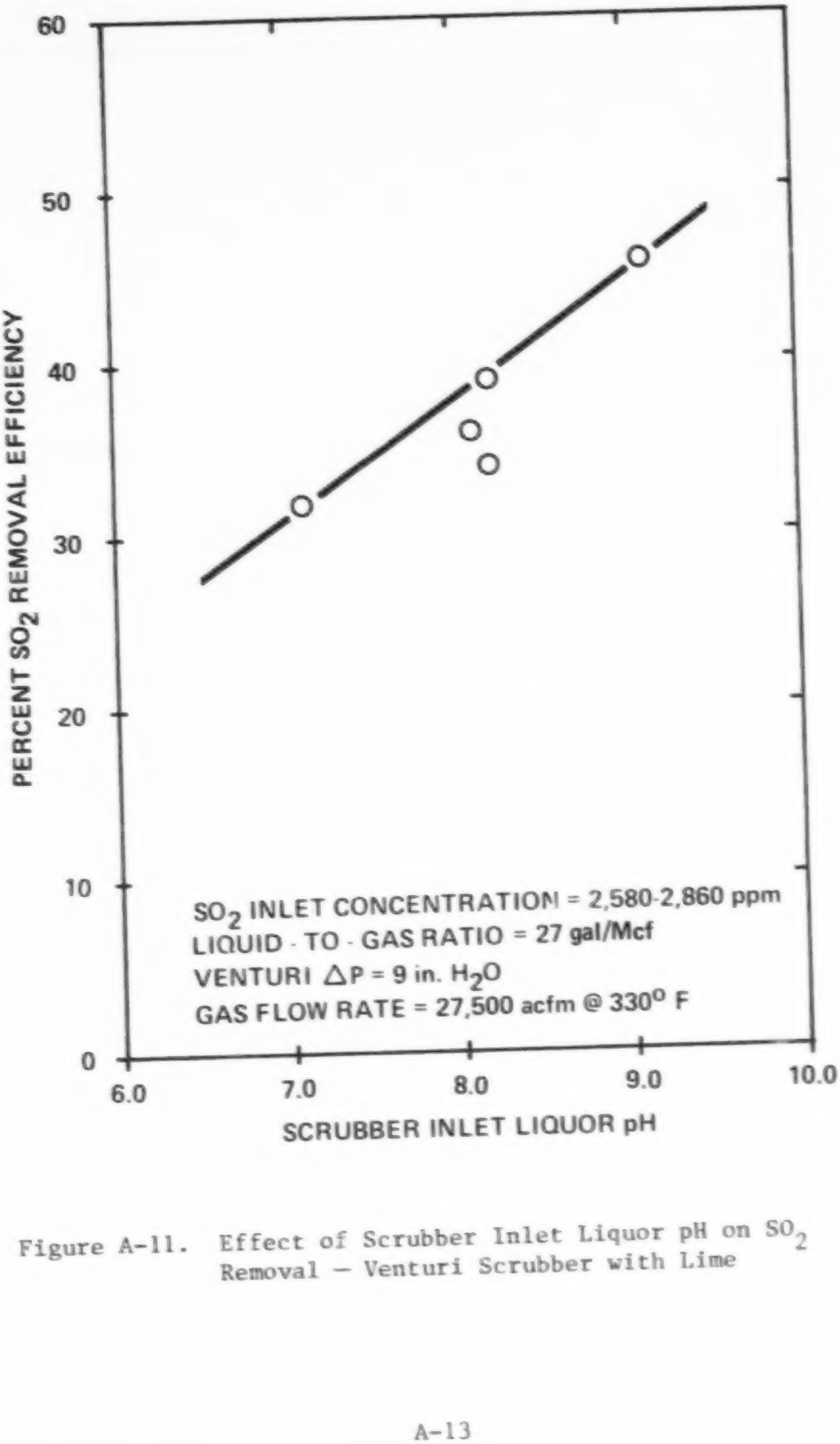

Figure A-11. Effect of Scrubber Inlet Liquor pH on SO_2 Removal — Venturi Scrubber with Lime

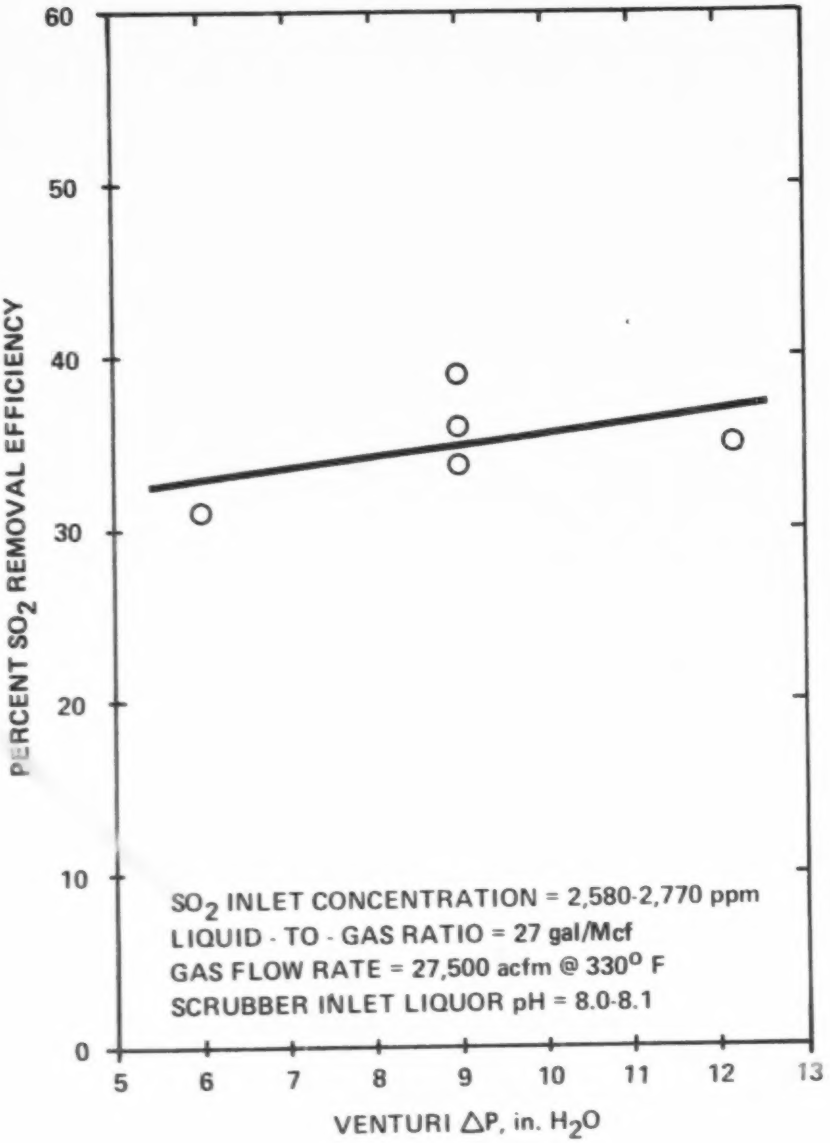

Figure A-12. Effect of Venturi ΔP on SO_2 Removal — Venturi Scrubber with Lime

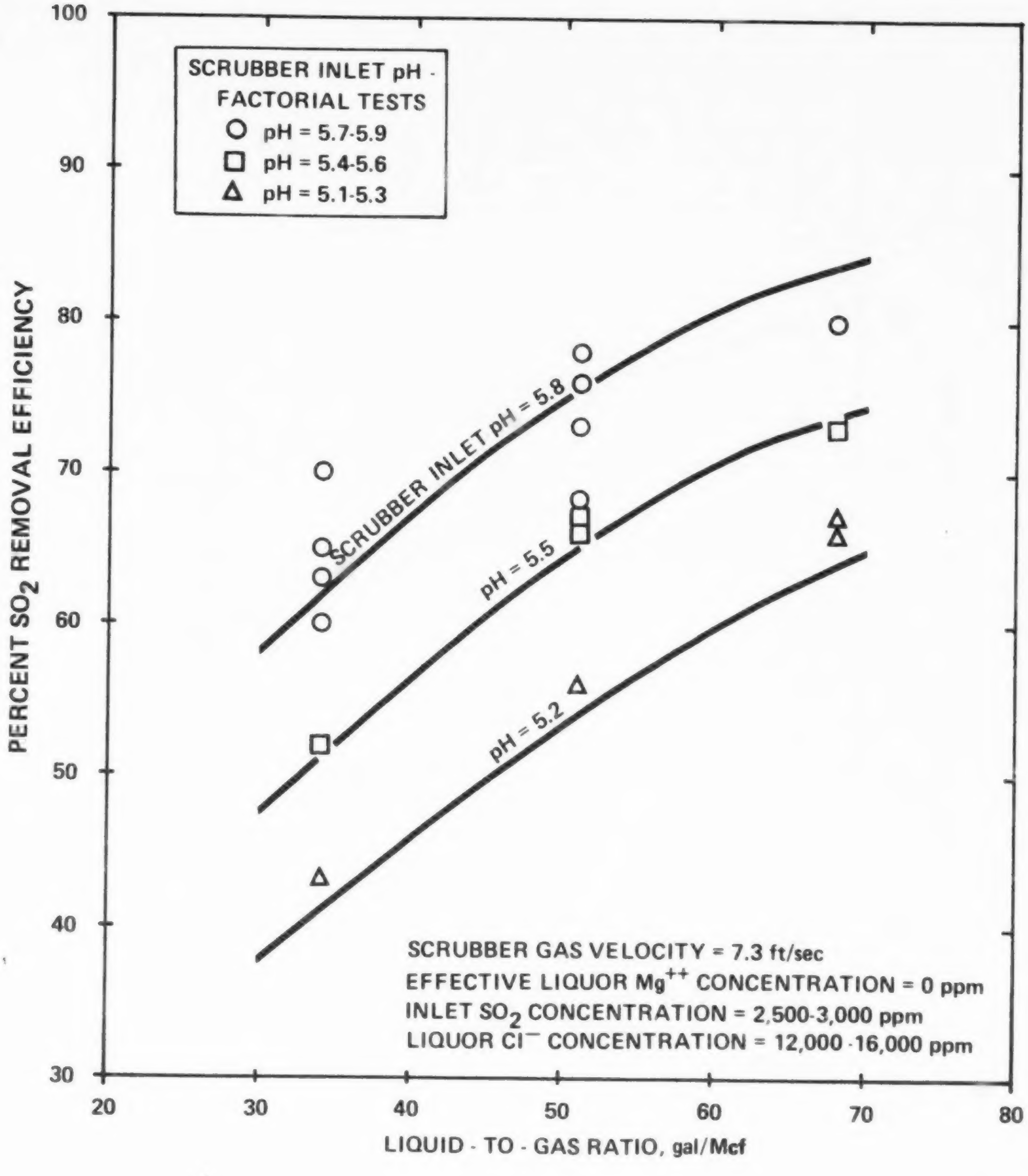

Figure A-13. Liquid-to-Gas Ratio and Scrubber Inlet pH Versus Predicted (Equation B-2) and Measured SO_2 Removal — Spray Tower with Limestone

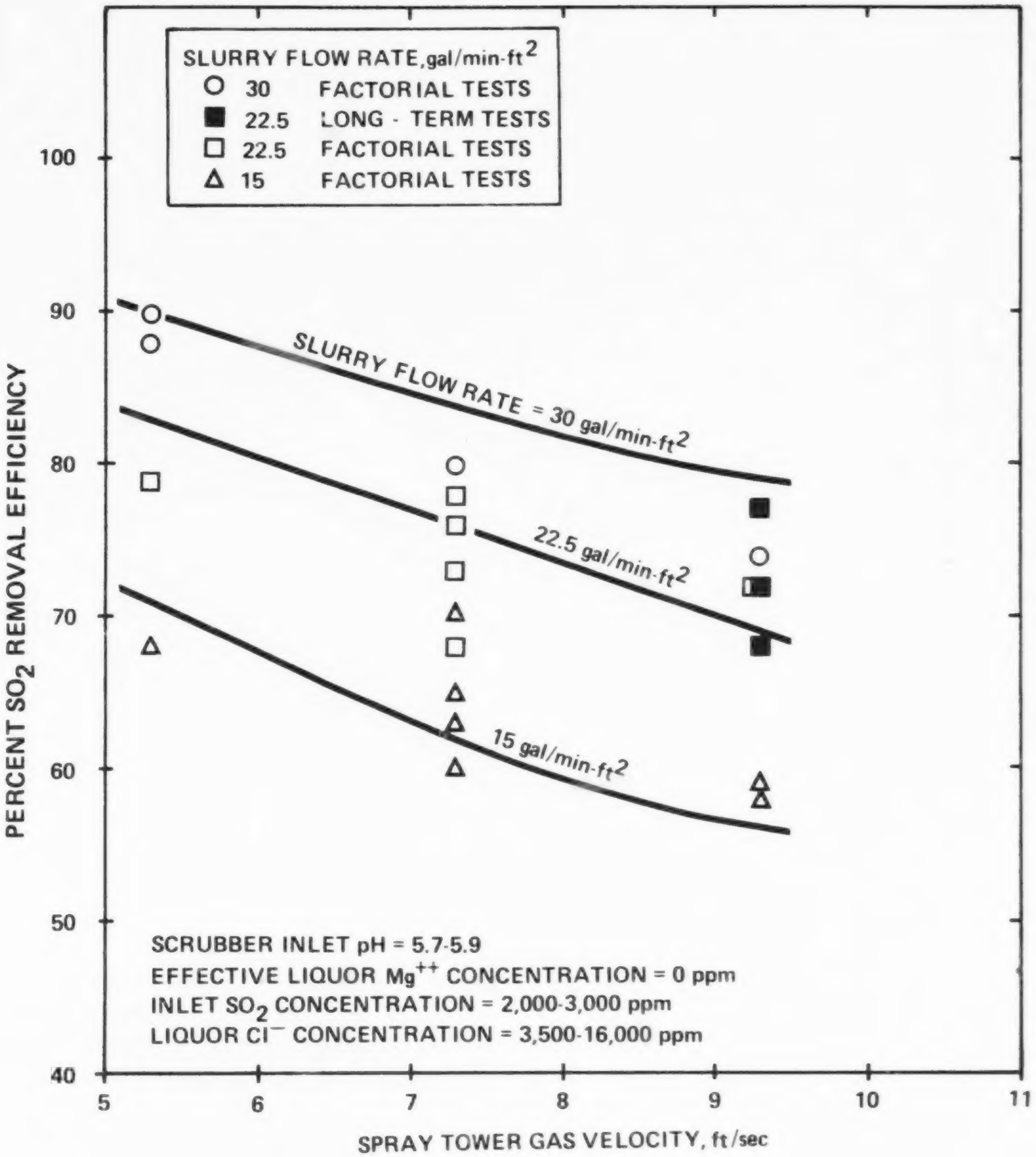

Figure A-14. Gas Velocity and Slurry Flow Rate Versus Predicted (Equation B-2) and Measured SO_2 Removal — Spray Tower with Limestone

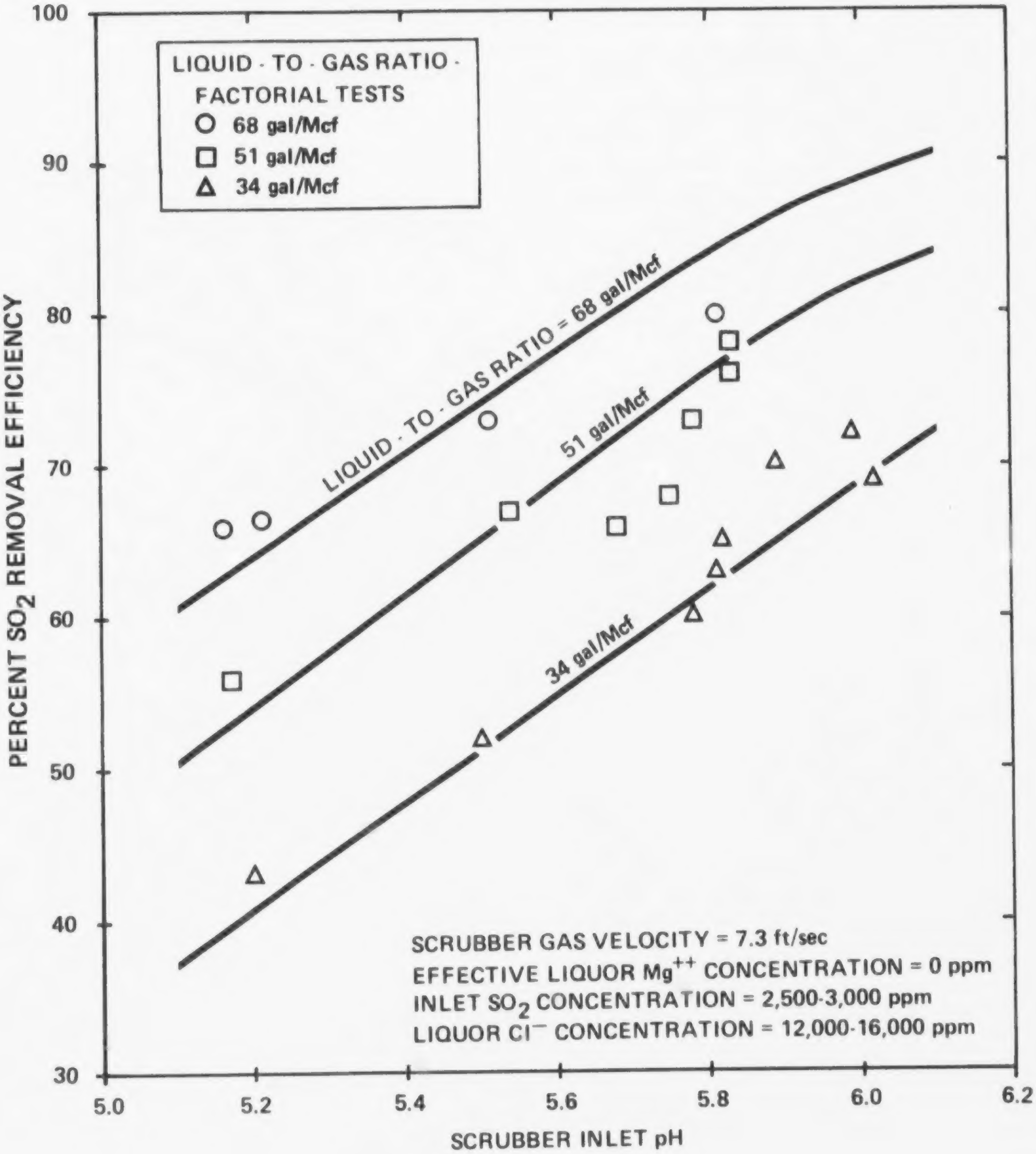

Figure A-15. Scrubber Inlet pH and Liquid-to-Gas Ratio Versus Predicted (Equation B-2) and Measured SO_2 Removal — Spray Tower with Limestone

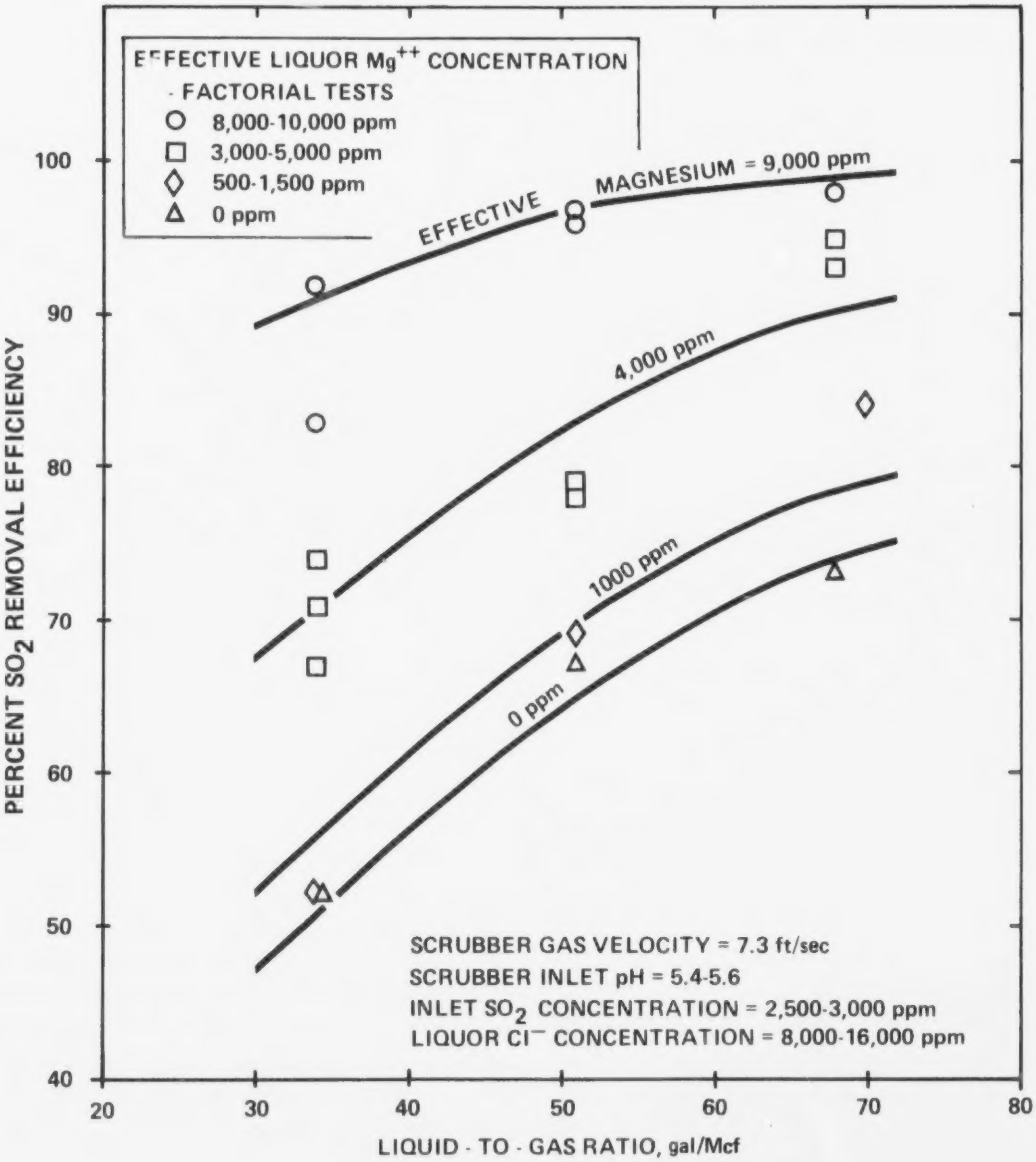

Figure A-16. Liquid-to-Gas Ratio and Effective Magnesium Versus Predicted (Equation B-2) and Measured SO_2 Removal — Spray Tower with Limestone

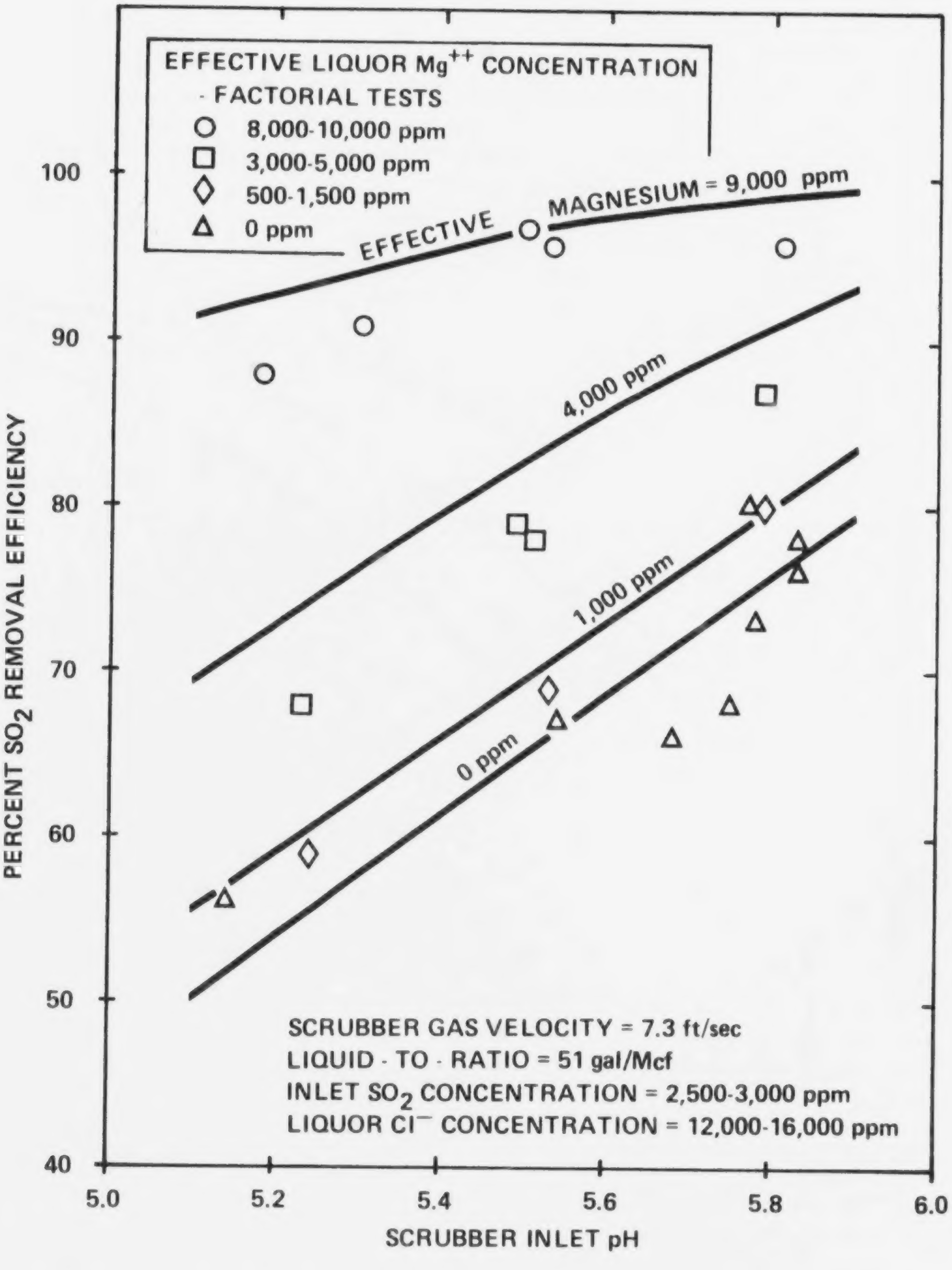

Figure A-17. Scrubber Inlet pH and Effective Magnesium Versus Predicted (Equation B-2) and Measured SO_2 Removal — Spray Tower with Limestone

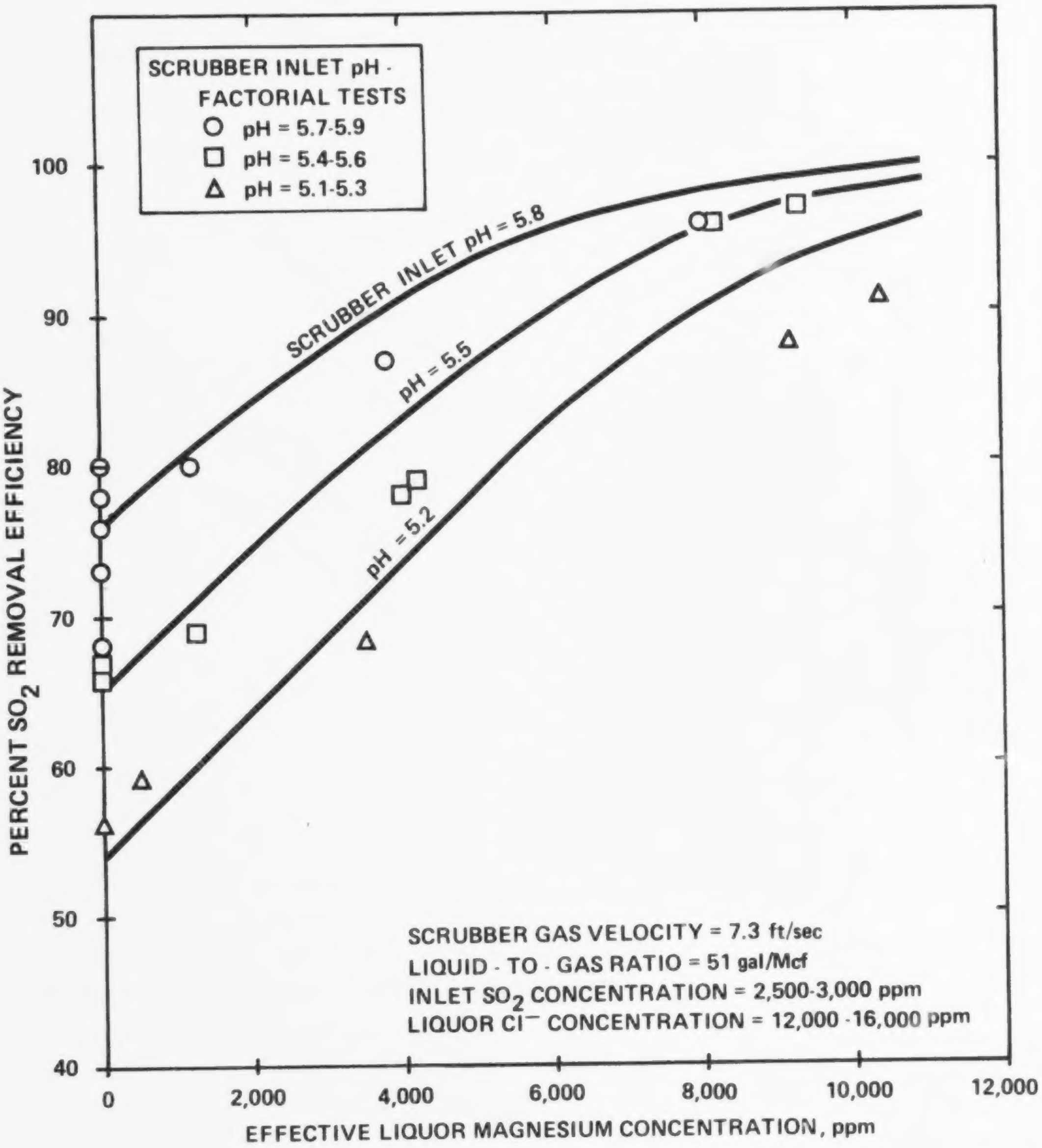

Figure A-18. Effective Magnesium and Scrubber Inlet pH Versus Predicted (Equation B-2) and Measured SO_2 Removal — Spray Tower with Limestone

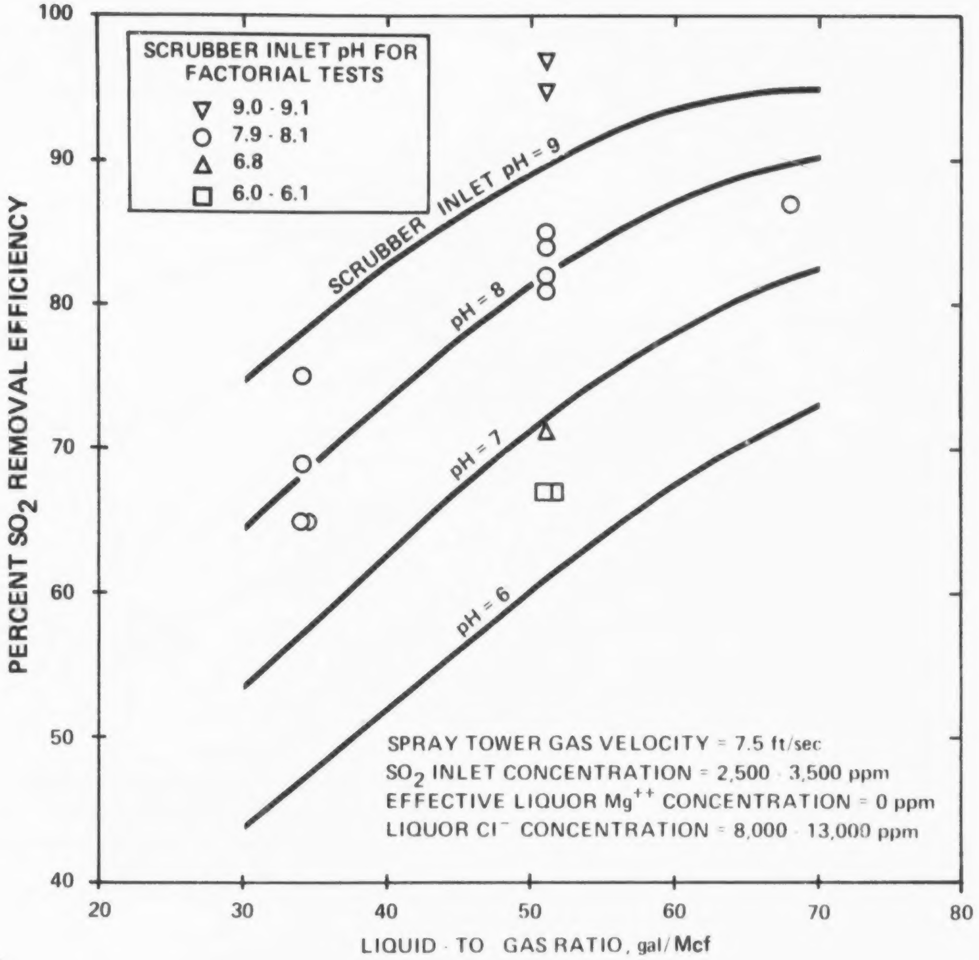

Figure A-19. Liquid-to-Gas Ratio and Scrubber Inlet pH Versus Predicted (Equation B-1) and Measured SO_2 Removal — Spray Tower with Lime

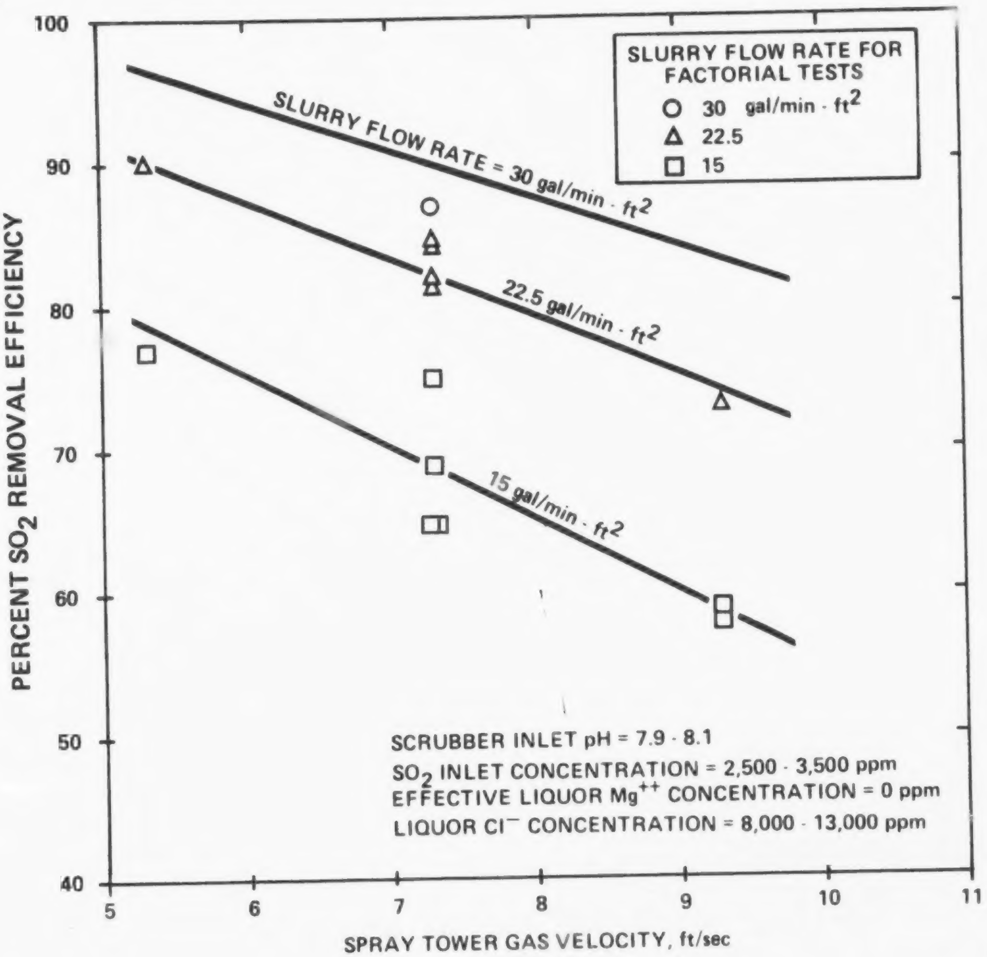

Figure A-20. Gas Velocity and Slurry Flow Rate Versus Predicted (Equation B-1) and Measured SO_2 Removal — Spray Tower with Lime

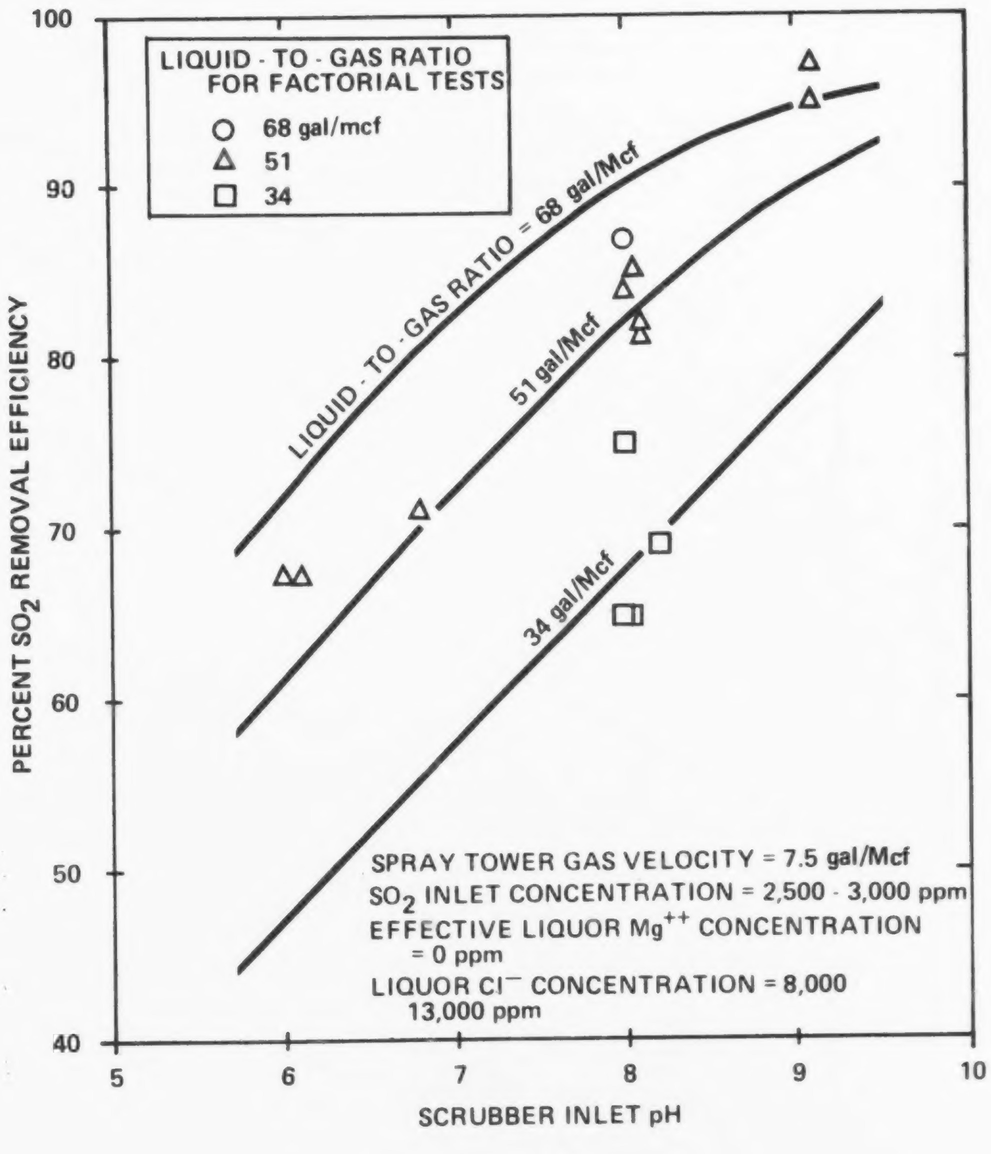

Figure A-21. Scrubber Inlet pH and Liquid-to-Gas Ratio Versus Predicted (Equation B-1) and Measured SO_2 Removal — Spray Tower with Lime

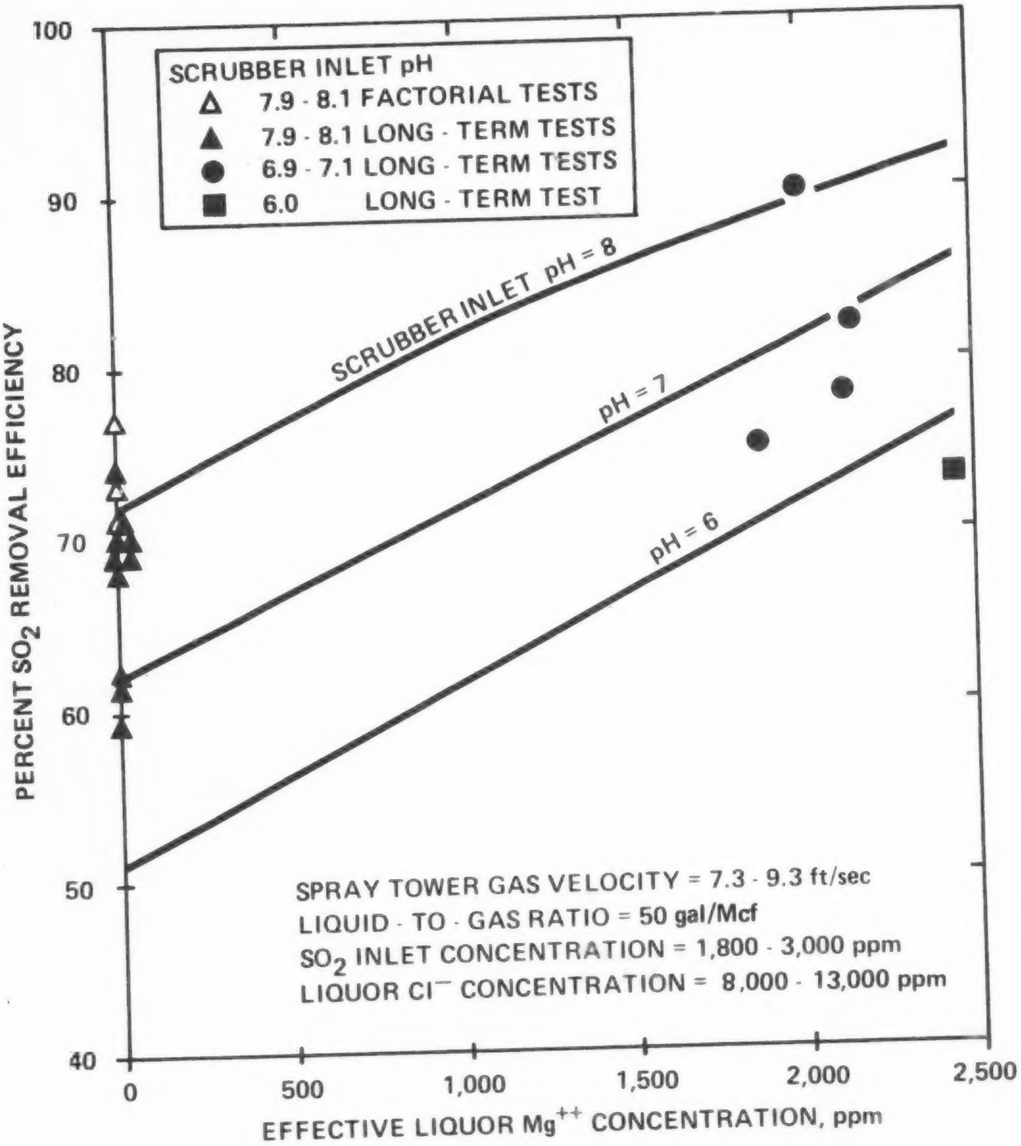

Figure A-22. Effective Magnesium and Scrubber Inlet pH Versus Predicted (Equation B-1) and Measured SO_2 Removal — Spray Tower with Lime

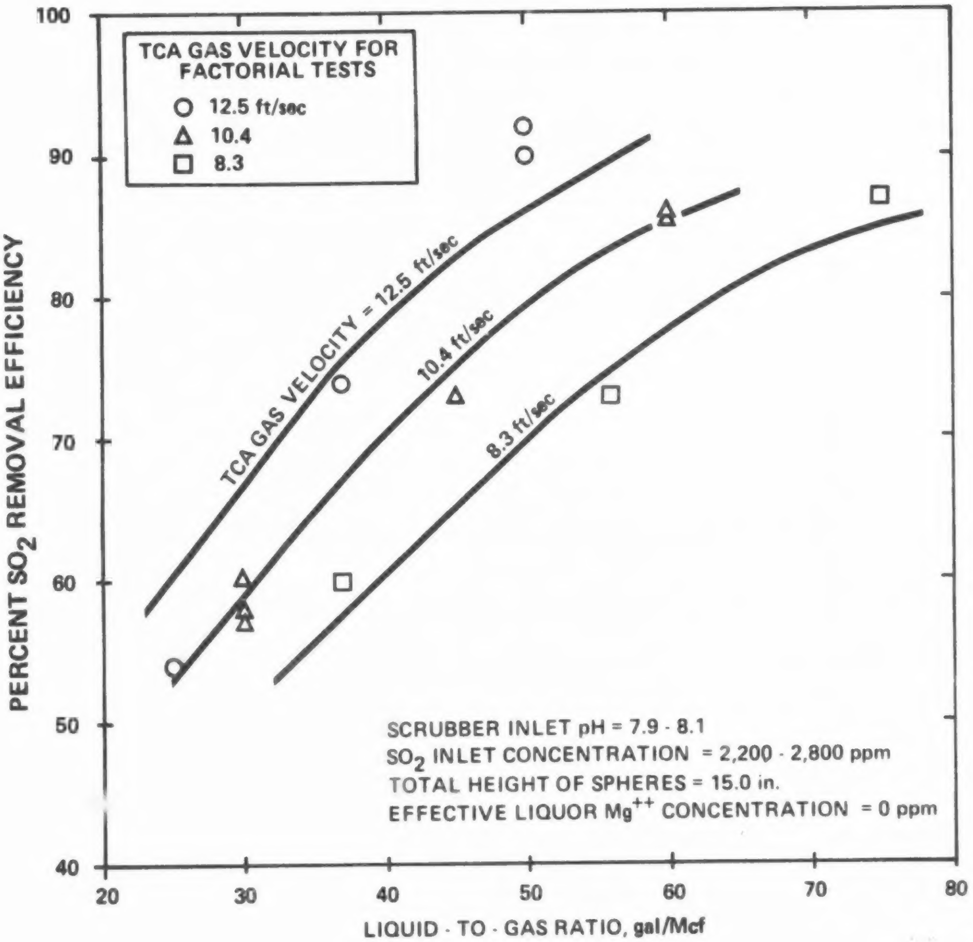

Figure A-23. Liquid-to-Gas Ratio and Scrubber Gas Velocity Versus Predicted (Equation B-3) and Measured SO_2 Removal — TCA with Lime

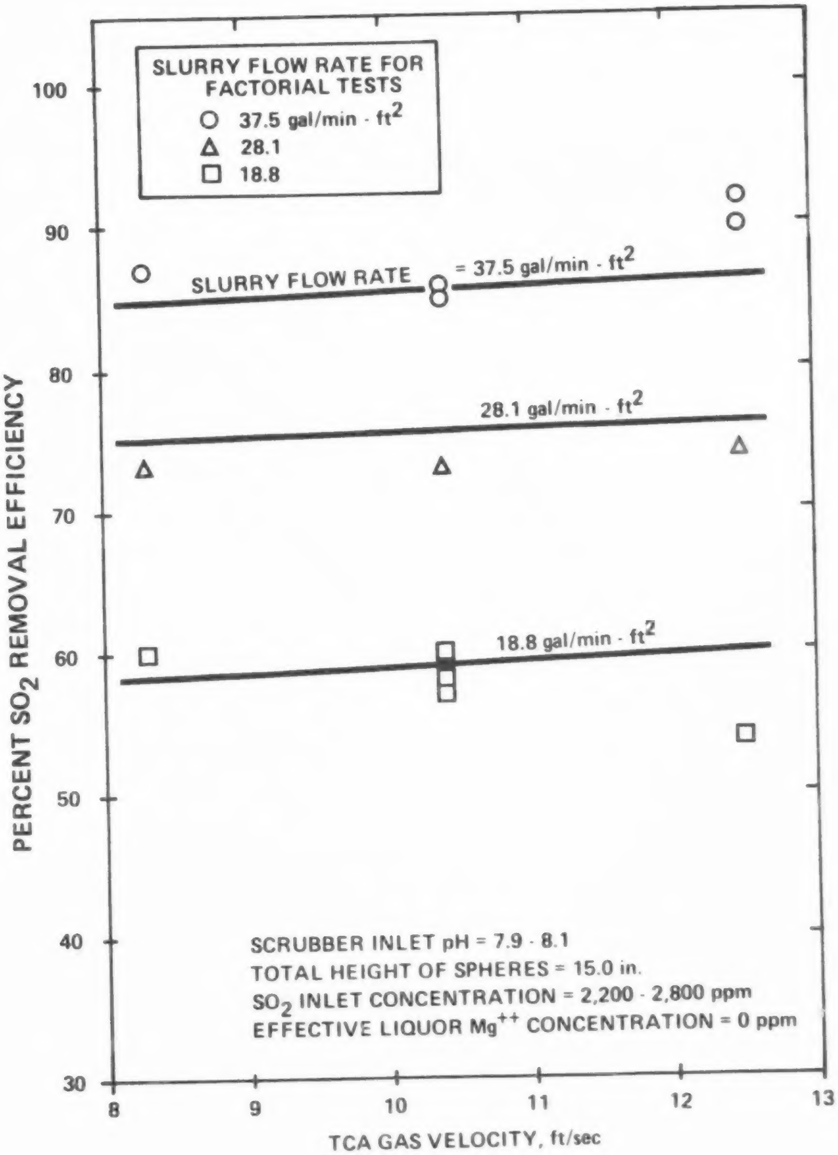

Figure A-24. TCA Gas Velocity and Slurry Flow Rate Versus Predicted (Equation B-3) and Measured SO_2 Removal — TCA with Lime

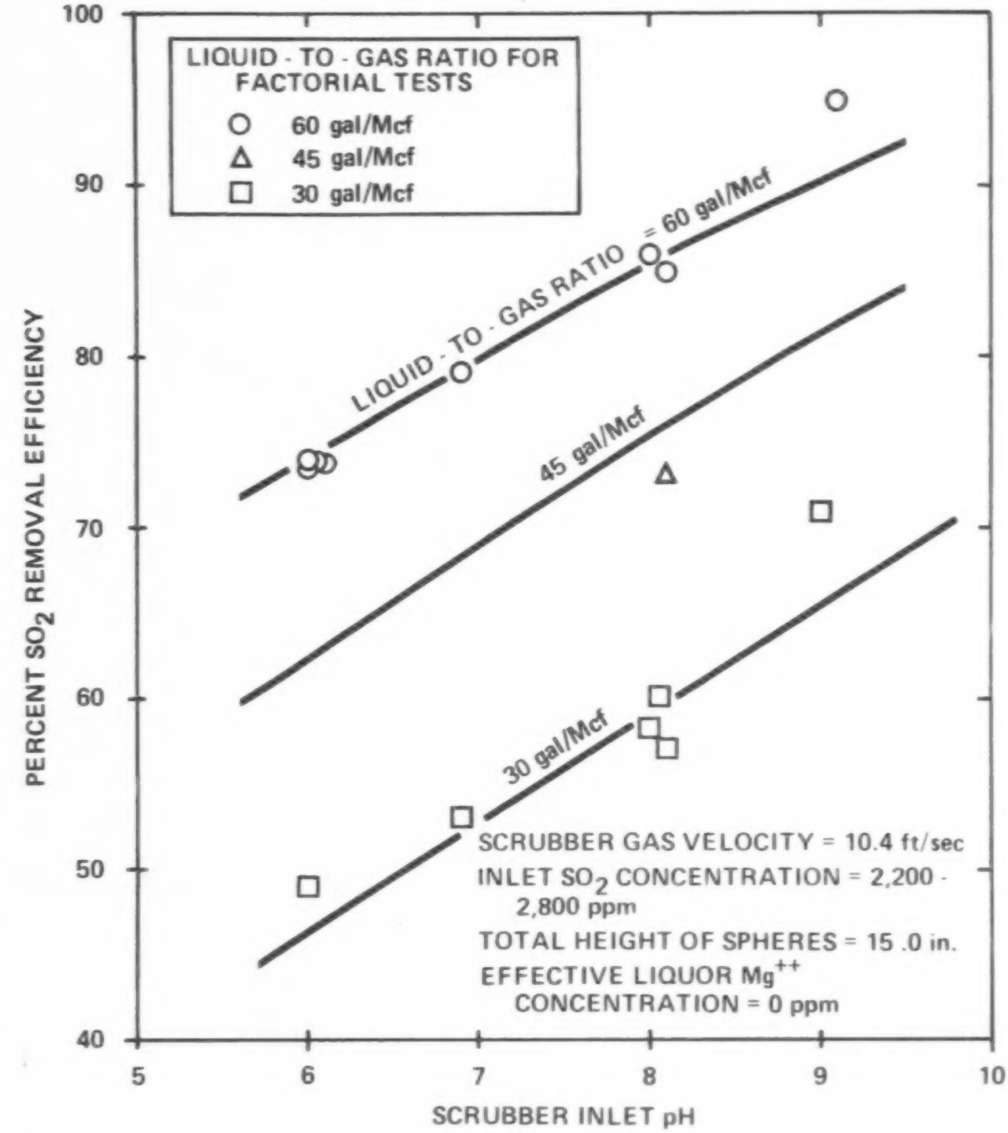

Figure A-25. Scrubber Inlet pH and Liquid-to-Gas Ratio Versus Predicted (Equation B-3) and Measured SO_2 Removal — TCA with Lime

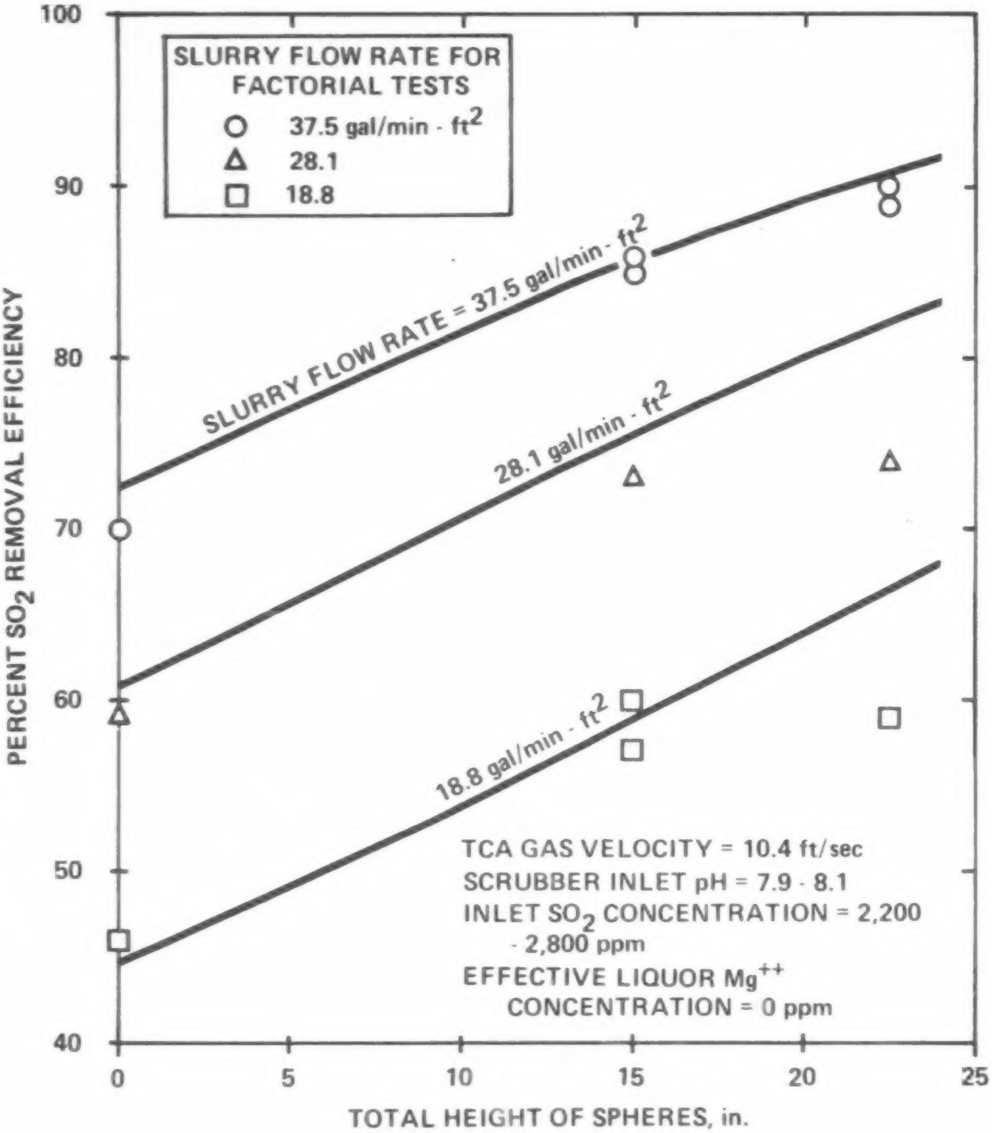

Figure A-26. Total Height of Spheres and Slurry Flow Rate Versus Predicted (Equation B-3) and Measured SO_2 Removal — TCA with Lime

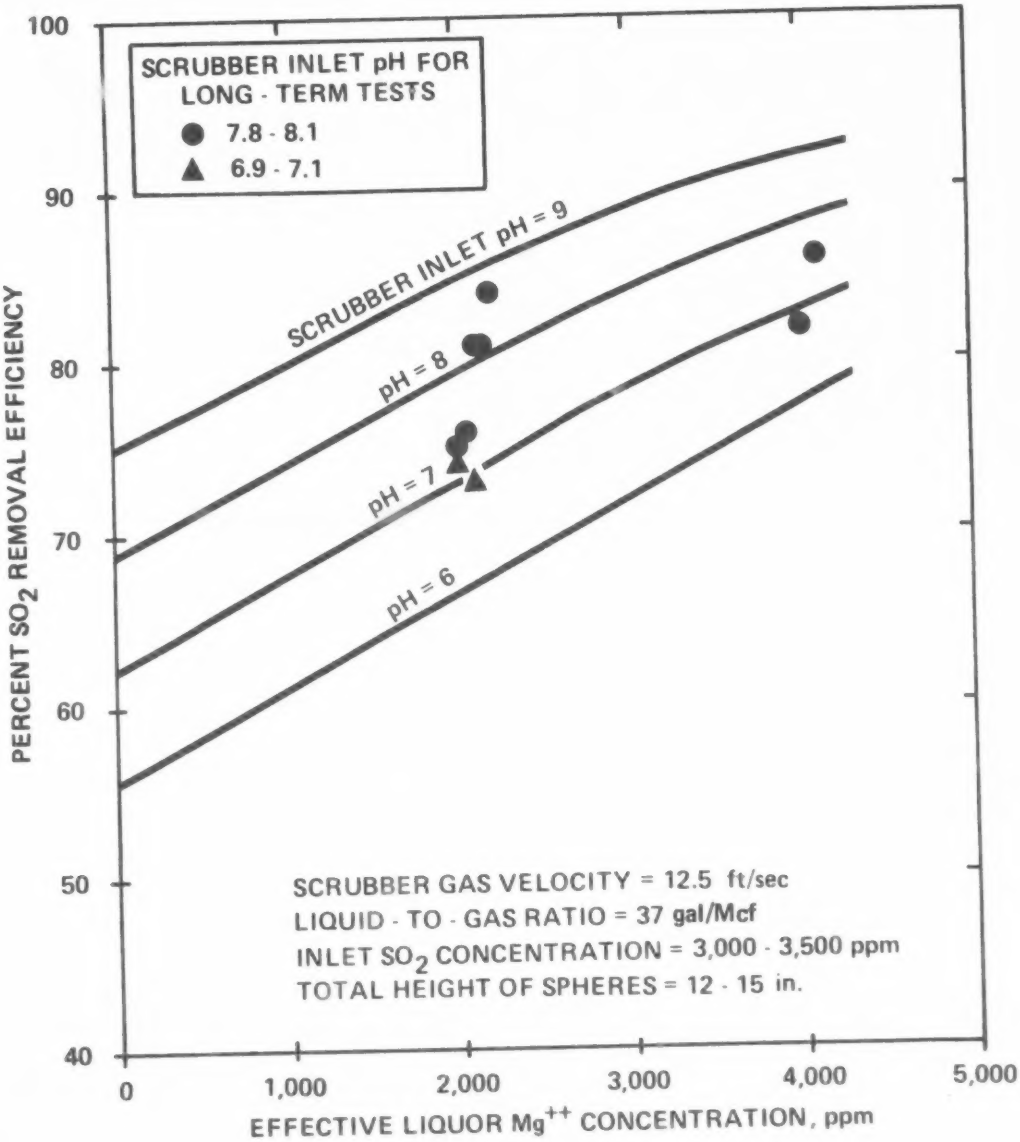

Figure A-27. Effective Magnesium and Scrubber Inlet pH Versus Predicted (Equation B-3) and Measured SO_2 Removal — TCA with Lime

APPENDIX B

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Appendix B

MATHEMATICAL MODELS AND NOMOGRAPHS FOR SO$_2$ REMOVAL BY LIMESTONE AND LIME WET SCRUBBING

A semitheoretical model has been developed and fitted to the Shawnee test data for predicting SO$_2$ removal by lime or limestone wet scrubbing from operating variables in a spray tower or a TCA. This model accounts for all major variables investigated at the Shawnee Test Facility except for possible effects of liquor sulfate (gypsum) saturation or sulfite oxidation on SO$_2$ removal at high magnesium-ion concentrations. Such effects are relatively insignificant at low magnesium-ion concentrations The fitted equations are as follows:

- Spray tower equation for SO$_2$ removal by lime slurry

$$\text{Fraction SO}_2\text{ Removal} = 1 - \exp\left\{- 0.0020\ (L/G)\ \exp\left[0.29\ pH_i + 2.8 \times 10^{-4}\ (Mg)_e + 4.7 \times 10^{-5}\ Cl\right]\right\} \qquad \text{(B-1)}$$

- Spray tower equation for SO$_2$ removal by limestone slurry

$$\text{Fraction SO}_2\text{ Removal} = 1 - \exp\left\{- 9.8 \times 10^{-5} (L/G)^{0.92}\, v^{0.19}\ \exp\left[pH_i + 1.35 \times 10^{-4} (Mg)_e - 1.7 \times 10^{-4} (SO_2)_i + 1.45 \times 10^{-5} Cl)\right]\right\} \qquad \text{(B-2)}$$

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- TCA equation for SO_2 removal by lime slurry

$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\Big\{- 0.0010\ (L/G)^{1.12}\ v^{0.65}$$

$$\exp\Big[0.18\ pH_i + 1.5 \times 10^{-4}(Mg)_e$$

$$- 2.2 \times 10^{-4}(SO_2)_i + 0.0039v\left(\frac{h_{tot}}{d_s} + N_G\right)\Big]\Big\} \quad \text{(B-3)}$$

- TCA equation for SO_2 removal by limestone slurry

$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\Big\{- 2.05 \times 10^{-4}(L/G)^{0.81}\ v^{0.36}$$

$$\exp\Big[4.3 \times 10^{-3}v\left(\frac{h_{tot}}{d_s} + N_G\right) + 0.81\ pH_i$$

$$+ 7.9 \times 10^{-5}\ (Mg)_e - 1.7 \times 10^{-4}(SO_2)_i$$

$$+ 1.3 \times 10^{-5}\ Cl\Big]\Big\} \qquad \text{(B-4)}$$

where:

Cl = measured total dissolved chloride concentration, ppm

d_s = diameter of TCA sphere, 1.5 inches at Shawnee

h_{tot} = total height of spheres in TCA, inches (zero for spray tower)

L/G = liquid-to-gas ratio in scrubber (125°F, humidified gas), gal/Mcf

$(Mg)_e$ = effective magnesium-ion concentration, ppm

$= [\text{ppm } Mg^{++} - (\text{ppm } Cl^{-}/2.92)]$ for $Mg^{++} > Cl^{-}/2.92$

$= 0$ for $Mg^{++} \leq Cl^{-}/2.92$

where 2.92 = ratio by weight of Cl^{-} to Mg^{++} in $MgCl_2$

N_G = number of grids or screens in TCA, 4 at Shawnee (zero for spray tower)

pH_i = scrubber inlet liquor pH

$(SO_2)_i$ = inlet gas SO_2 concentration, ppm

v = gas velocity in scrubber (125°F, humidified gas), ft/sec

These equations are verified only for the range of conditions encountered at the Shawnee Test Facility. Extrapolation significantly beyond this range is not recommended. The ranges of the important operating variables are given below:

Equation B-1

L/G = 30 to 100 gal/Mcf
pH_i = 6.0 to 9.0
$(Mg)_e$ = 0. to 2,500 ppm
Cl = 0 to 12,000 ppm
v = 5.3 to 9.3 ft/sec
$(SO_2)_i$ = 2,000 to 4,000 ppm

Equation B-2

L/G = 30 to 70 gal/Mcf
v = 5.3 to 9.3 ft/sec
pH_i = 5.2 to 5.9
$(Mg)_e$ = 0 to 10,000 ppm
$(SO_2)_i$ = 1,500 to 4,500 ppm
Cl = 4,000 to 16,000 ppm

Equation B-3

L/G = 25 to 75 gal/Mcf
v = 8.5 to 12.5 ft/sec
h_{tot} = 0 to 22.5 inches
d_s = 1.5 inches
N_G = 4
pH_i = 6.0 to 9.0
$(Mg)_e$ = 0 to 4,000 ppm
$(SO_2)_i$ = 2,000 to 4,000 ppm
Cl = 0 to 8,000 ppm

Equation B-4

L/G = 25 to 75 gal/Mcf
v = 8.5 to 12.5 ft/sec
pH_i = 5.1 to 5.9
$(Mg)_e$ = 0 to 10,000 ppm
$(SO_2)_i$ = 1,500 to 4,500 ppm
Cl = 1,500 to 20,000 ppm
d_s = 1.5 inches
h_{tot} = 0 to 22.5 inches
N_G = 4

Figures B-1 through B-5 are nomographs for predicting, from operating variables, SO_2 removal by limestone or lime wet scrubbing with a spray tower or a TCA. Figure B-1 can be used to predict the effect of a change in SO_2 inlet concentration on SO_2 removal. Nomographs for lime and limestone scrubbing with a spray tower, Figures B-2 and B-3, respectively, are derived from Equations B-1 and B-2. Figures B-4 and B-5 for lime and limestone scrubbing with a TCA, respectively, are derived from Equations B-3 and B-4. The nomographs cover the same ranges of operation as the corresponding equations

In practice, predicting SO_2 removal by nomographs is typically within about one-half percent of the corresponding equation prediction. Speed of calculation using one of these nomographs is about the same as for solving the corresponding equation by electronic calculator.

The inset in Figure B-2 illustrates the use of these nomographs. First, the liquid-to-gas ratio is located on the bottom curve. Then, a vertical line is drawn to intersect with the next lowest horizontal line, which represents the lowest level of the next variable — scrubber inlet liquor pH in this case. A line is then drawn upward from the horizontal line and parallel to the nearest pH curve, until the y-axis level corresponding to the specified pH, 8.0 in this case, is reached. Again, a vertical line is drawn to the horizontal line for the next operating variable chloride-ion concentration. This procedure is continued until the top of the nomograph is reached. The predicted SO_2 removal is then the x-axis value at this point. In the example case, the nomograph SO_2 removal prediction by Figure B-2 was 85 percent; Equation B-1 correspondingly predicted 86 percent.

If SO_2 removal is specified in a design, and prediction of another operating variable is required, for example scrubber inlet liquor pH, this prediction can be made from the nomograph by working upwards from the bottom of the nomograph and downwards from the top. The two resultant intersecting lines, one parallel to the nomograph pH curves from the bottom up and the other vertical from the top of the pH scale, meet at the y-axis value of the operating pH.

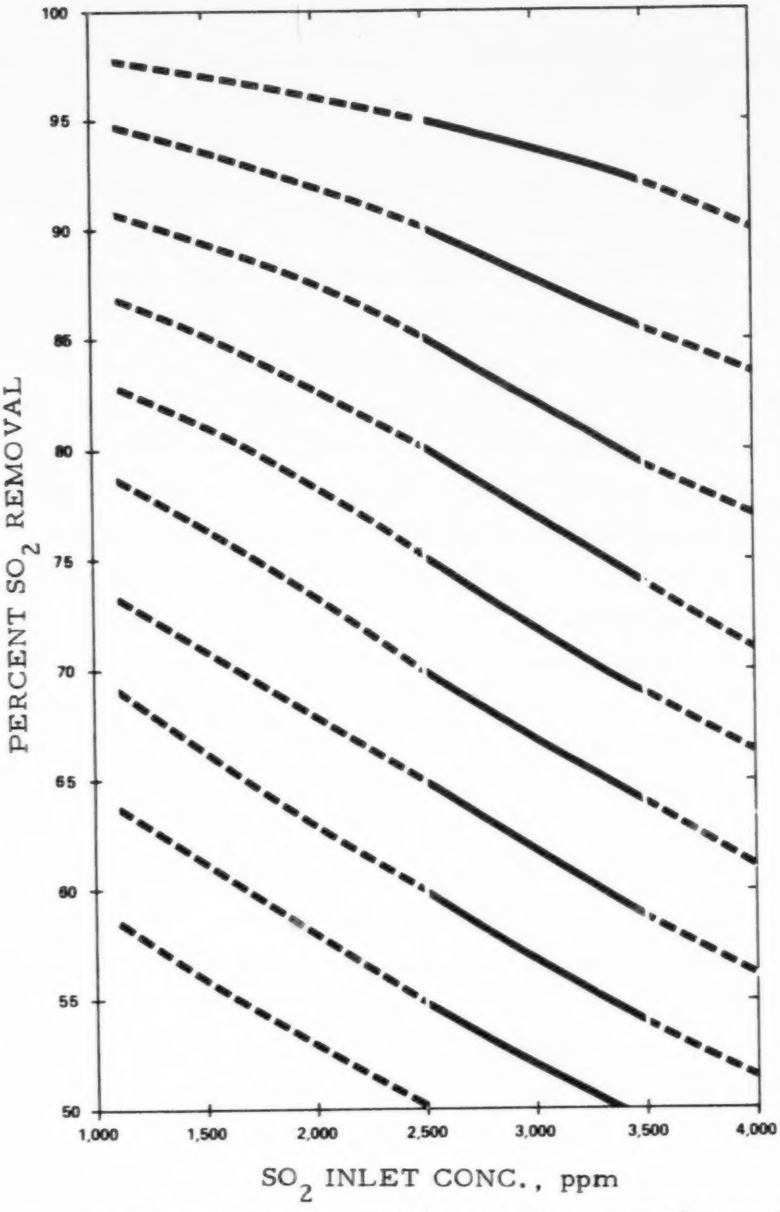

Figure B-1. Predicted Effect of Inlet SO_2 Concentration on SO_2 Removal

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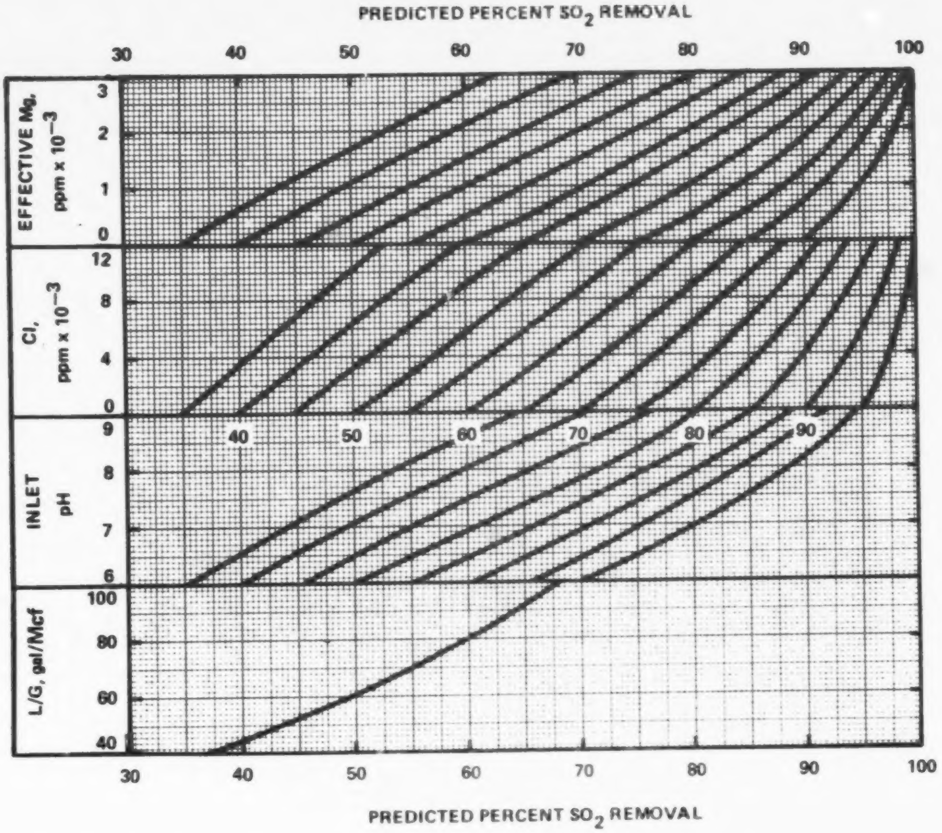
Figure B-1 can be used to predict the effect of a change in SO_2 inlet concentration on SO_2 removal. When the SO_2 removal, $(SO_2R)1$, at a particular inlet SO_2 concentration, $(SO_2i)1$, is known, the removal, $(SO_2R)2$, at a different inlet SO_2 concentration, $(SO_2i)2$, is obtained by first locating the point $[(SO_2i)1, (SO_2R)1]$ and then paralleling the nearest curve until $(SO_2i)2$ is reached. The corresponding value on the y axis is then $(SO_2R)2$.



$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\left\{-0.0020\,(L/G)\exp\left[\,0.29\,pH_i + 2.8\times10^{-4}\,(Mg)_e + 4.7\times10^{-5}\,Cl\right]\right\}$$

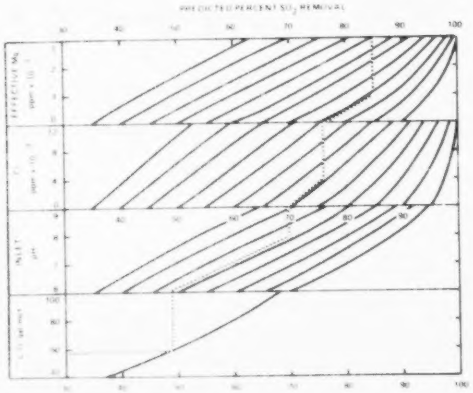
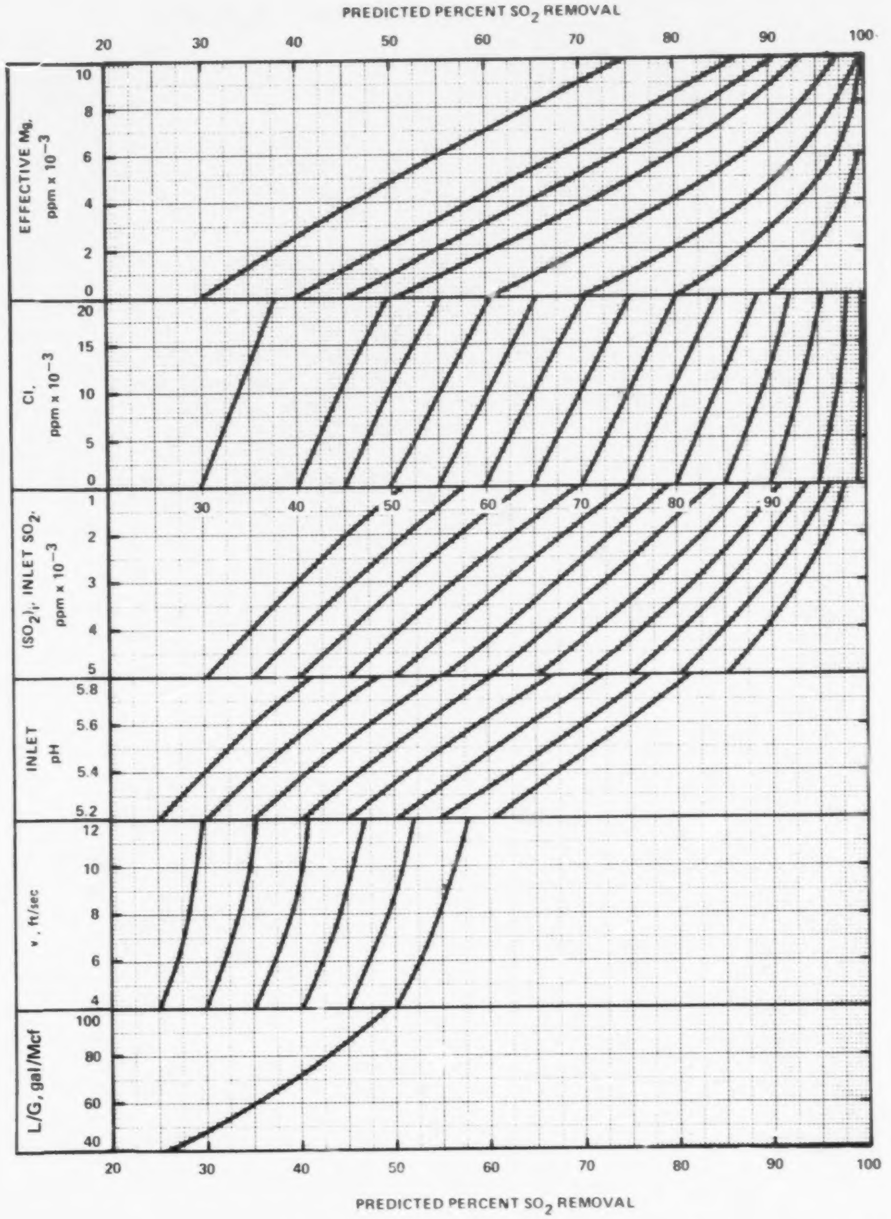
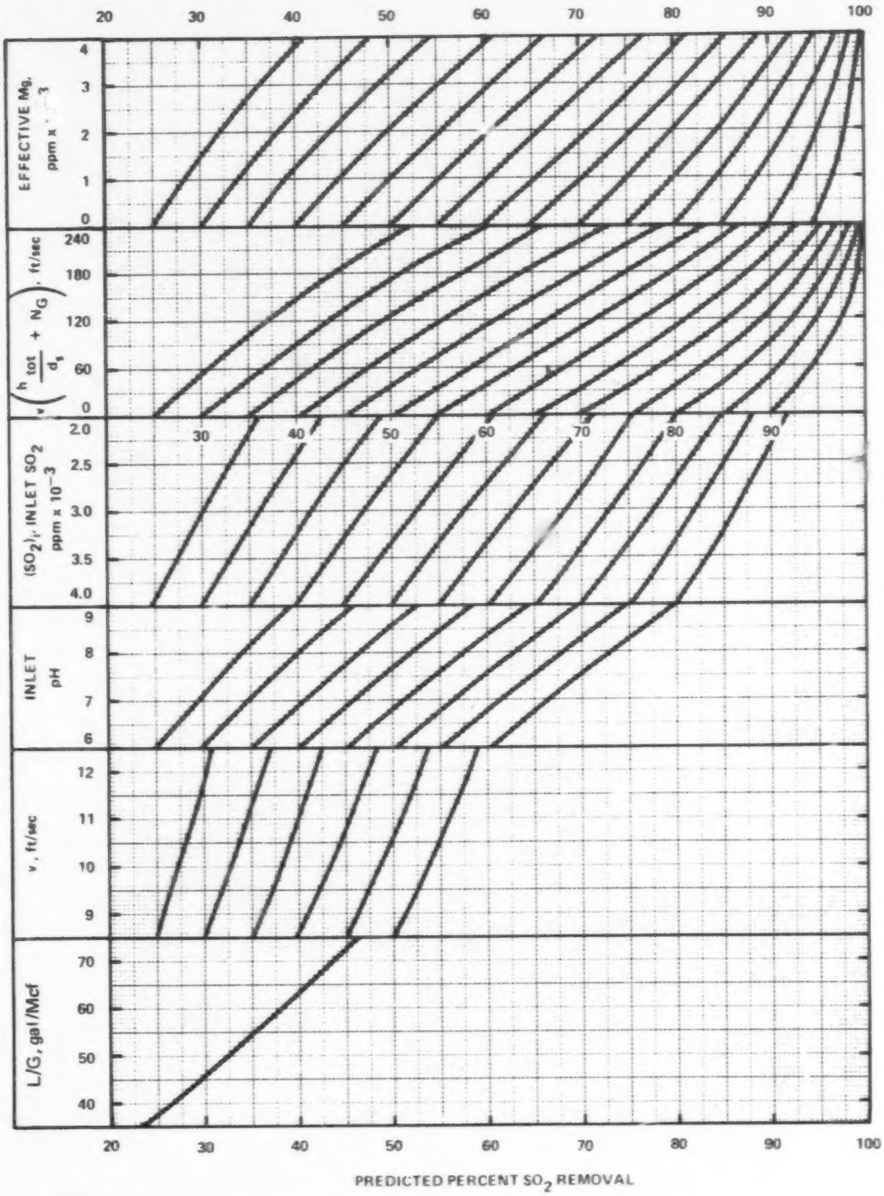


Figure B-2. Nomograph for Percent SO_2 Removal by Lime Scrubbing in a Spray Tower



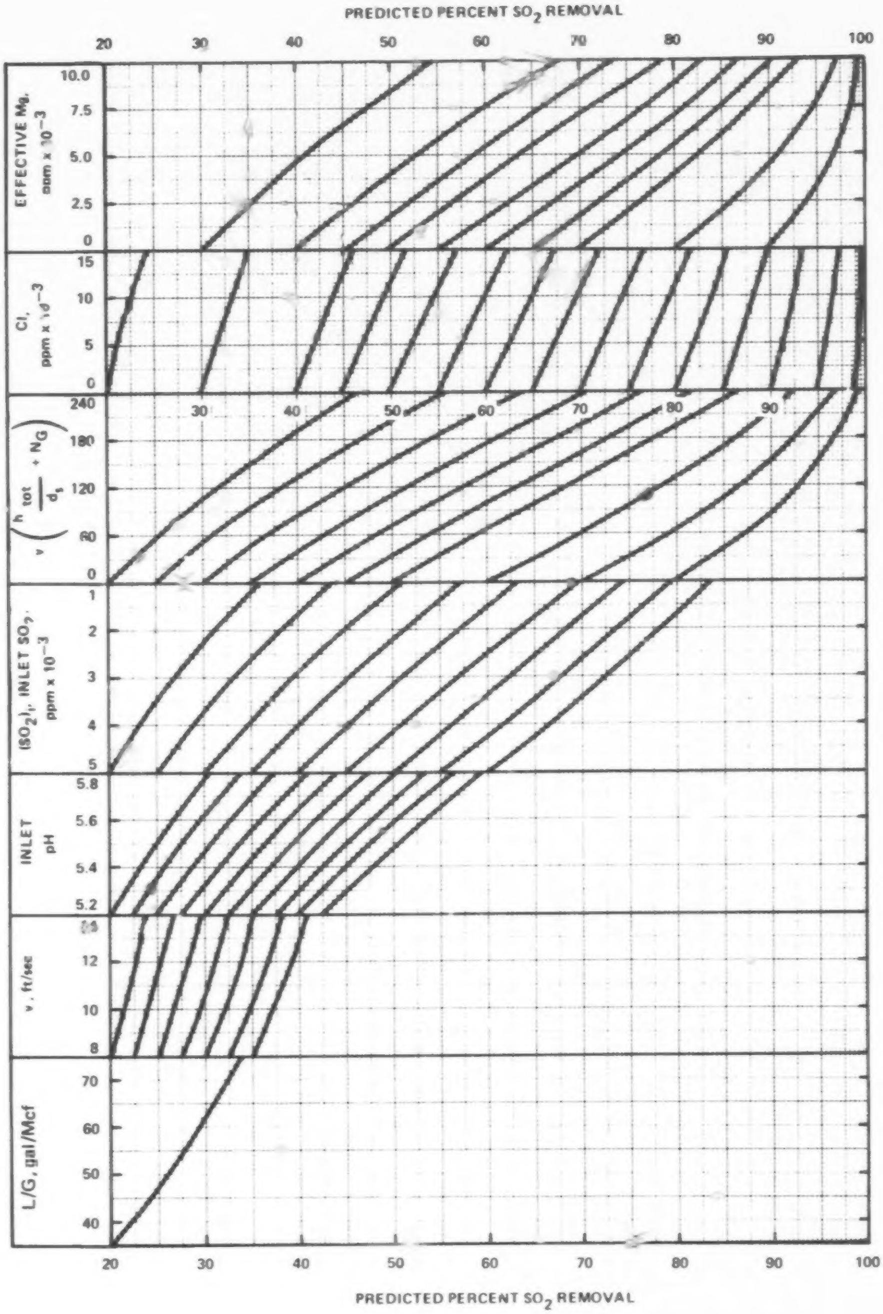
$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\left\{-9.8 \times 10^{-5} (L/G)^{0.92} v^{0.19} \exp\left[pH_i + 1.35 \times 10^{-4} (Mg)_e - 1.7 \times 10^{-4} v_i + 1.45 \times 10^{-5} Cl\right]\right\}$$

Figure B-3. Nomograph for Percent SO_2 Removal by Limestone Scrubbing in a Spray Tower



$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\left\{-0.0010\,(L/G)^{1.12}\,v^{0.65}\,\exp\left[0.18\,pH_i + 1.5 \times 10^{-4}\,(Mg)_e - 2.2 \times 10^{-4}\,y_i + 0.0039\,v\left(\frac{hp}{dp} + N_G\right)\right]\right\}$$

Figure B-4. Nomograph for Percent SO_2 Removal by Lime Scrubbing in a TCA



$$\text{Fraction } SO_2 \text{ Removal} = 1 - \exp\left\{ -2.05 \times 10^{-4} (L/G)^{0.81} v^{0.36} \exp\left[4.3 \times 10^{-3} v\left(\frac{hp}{dp} + N_G\right) + 0.81\, pH_i + 7.9 \times 10^{-5} (Mg)_e - 1.7 \times 10^{-4} y_i + 1.3 \times 10^{-5}\, Cl\right]\right\}$$

Figure B-5. Nomograph for Percent SO2 Removal by Limestone Scrubbing in a TCA

Appendix C

CONVERSION TABLE

To Convert From	To	Multiply By
scfm ($60^{o}F$)	nm^3/hr ($0^{o}C$)	1.61
cfm	m^3/hr	1.70
^{o}F	^{o}C	(^{o}F-32)/1.8
ft	m	0.305
ft/hr	m/hr	0.305
ft/sec	m/sec	0.305
ft^2	m^2	0.0929
ft^2/tons per day	m^2/metric tons per day	0.102
gal/Mcf	1 m^3	0.134
gpm	1/min	3.79
gpm/ft^2	1/min/m^2	40.8
gr/scf	gm/m^3	2.29
in.	cm	2.54
in. H_2O	mm Hg	1.87
lb	gm	454
lb-moles	gm-moles	454
lb-moles/hr	gm-moles/min	7.56
lb-moles/hr ft^2	gm-moles/min/m^2	81.4
lb-moles/min	gm-moles/sec	7.56

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Plant Name and Company	Process	Vendor	New or Retrofit	Rating (MW)	% Sulfur in Coal	Fly A Collect
Cholla 1 Arizona Public Service	Limestone	Research Cottrell	R	115	0.4-1	–
Duck Creek 1A Central Illinois Light	Limestone	Riley Stoker–Environeering	N	100	2.5-3	ESP
La Cygne 1 Kansas City Power & Light	Limestone	Babcock & Wilcox	N	820	3-6	–
Lawrence 4 Kansas Power & Light	Limestone(c)	Combustion Engineering	R	125	0.5	–
Lawrence 5 Kansas Power & Light	Limestone(c)	Combustion Engineering	N	400	0.5	–
Martin Lake 1 Texas Utilities	Limestone	Research Cottrell	N	793	1-2 (lignite)	ESP
Sherburne 1 Northern States Power	Limestone	Combustion Engineering	N	710	0.8	–
Sherburne 2 Northern States Power	Limestone	Combustion Engineering	N	680	0.8	–
Southwest 1 Springfield City Utilities	Limestone	UOP	N	200	3.5	ESP
Widows Creek 8 TVA	Limestone	TVA	R	550	3.7	ESP
Will County 1 Commonwealth Edison	Limestone	Babcock & Wilcox	R	167	4	ESP
Bruce Mansfield 1 Pennsylvania Power	Lime with MgO	Chemico	N	835	4.7	–
Cane Run 4 Louisville Gas & Electric	Lime	American Air Filter	R	178	3.5-4	ES
Conesville 5 Columbus & Southern Ohio Elec.	Lime with MgO	UOP	N	400	4.5-4.9	ES

(a) Net removal is about 60% for two units.
(b) Shakedown operation only. Plant will burn low sulfur coal until 3 additional modules are built.
(c) Converted from boiler injection to conventional scrubbing–unit 4 in 1976, Unit 5 in 1977.
(d) Bypass reheat reduces net removal to about 60%.

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Table 3-1

OPERATIONAL FGD SYSTEMS ON U.S. COAL-FIRED UTILITY BOILERS

w or rofit	Rating (MW)	% Sulfur in Coal	Fly Ash Collector	SO2 Absorber	No. of Modules	L/G (gal/Mcf)	Design Efficiency %	Startup	Reference
	115	0.4-1	–	A-Venturi and packed tower B-Venturi	2	20 & 25 20	> 90(a) 45	10/73	1,2
	100	2.5-3	ESP	Ventri rod absorber	1	50	75-85	9/76(b)	1,3
	820	3-6	–	Venturi and sieve tray tower	7	33	60-80	2/73	1,3
	125	0.5	–	Marble bed—converted to venturi rod/spray tower in 1976	2	–	75-85	12/63	1
	400	0.5	–	Marble bed—being converted to venturi rod/spray tower in 1977	2	–	–	11/71	1
	793	1-2 (lignite)	ESP	Spray quencher/wetted film contactor	6	80	97(d)	4/77	2
	710	0.8	–	Venturi rod/marble bed	12	27	50-60	3/76	1,3
	680	0.8	–	Venturi rod/marble bed	12	–	50-60	4/77	1
	200	3.5	ESP	TCA	2	60	80-87	3/77	1,4
	550	3.7	ESP	Venturi/spray tower	4	50	80	5/77	1
	167	4	ESP	Venturi/two-stage perforated plate tower	2	34	85	2/72	3
	835	4.7	–	2-stage venturi	6	–	92	4/76	1
	178	3.5-4	ESP	Mobile bed	2	30-40	90	8/76	1,3
	400	4.5-4.9	ESP	TCA	2	60	90	3/77	1,4

re built.
77.

Plant Name and Company	Process	Vendor	New or Retrofit	Rating (MW)	% Sulfur in Coal	Fly Ash Collector	SO
Elrama Duquesne Light	Lime with MgO	Chemico	R	510	1-2.8	ESP	Ventur
Green River 1,2,3 Kentucky Utilities	Lime	American Air Filter	R	64	3.8-4	Mechanical	Ventur
Hawthorn 3 Kansas City Power & Light	Lime(a)	Combustion Engineering	R	140	0.5-3.5	–	Marble
Hawthorn 4 Kansas City Power & Light	Lime(a)	Combustion Engineering	R	100	0.5-3.5	–	Marble
Paddys Run 6 Louisville Gas & Electric	Lime (Carbide)	Combustion Engineering	R	65	3.5-4	ESP	Marble
Phillips Duquesne Light	Lime with MgO	Chemico	R	410	1-2.8	ESP	Three tra One 2- tra
EPA Shawnee Test Facility Tennessee Valley Authority	Lime and limestone	Chemico UOP	R	10 10	3-5	(b)	Ventur TCA
Colstrip 1 Montana Power	Alkaline fly ash with lime	ADL/CEA	N	360	0.8	–	Ventur
Colstrip 2 Montana Power	Alkaline fly ash with lime	ADL/CEA	N	360	0.8	–	Ventur
Reid Gardner 1 Nevada Power	Sodium carbonate	ADL/CEA	R	125	0.5-1	Mechanical	Ventu
Reid Gardner 2 Nevada Power	Sodium carbonate	ADL/CEA	R	125	0.5-1	Mechanical	Ventu
Reid Gardner 3 Nevada Power	Sodium carbonate	ADL/CEA	N	125	0.5-1	Mechanical	Ventur
Eddystone 1A Philadelphia Electric	Magnesium oxide	United Engineers–Peco	R	120	2.5	ESP/ Venturi	Ventu
D.H. Mitchell II Northern Indiana Public Service	Sodium sulfite (Wellman Lord)	Davy Powergas–Allied Chemical	R	115	3.2-3.5	ESP/ Venturi	Absor

(a) Converted from limestone boiler injection to conventional scrubbing in 1977
(b) Tests have been made with and without ESP

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Table 3-1 (Continued)

ndor	New or Retrofit	Rating (MW)	% Sulfur in Coal	Fly Ash Collector	SO_2 Absorber	No. of Modules	L/G (gal/mcf)	Design Efficiency %	Startup	Reference
	R	510	1-2.8	ESP	Venturi	5	30-35	85	10/75	3
Air Filter	R	64	3.8-4	Mechanical	Venturi/mobile bed	1	35	>80	9/75	1,3
n Engineering	R	140	0.5-3.5	–	Marble bed	2	–	70	11/72	1
n Engineering	R	100	0.5-3.5	–	Marble bed	2	–	70	8/72	1
n Engineering	R	65	3.5-4	ESP	Marble bed	2	–	80	4/73	1
	R	410	1-2.8	ESP	Three 1-stage venturi trains One 2-stage venturi train	4	30	83 90	7/73	1
	R	10 10	3-5	(b)	Venturi/spray tower TCA	1 1	Varies with each test		4/72	1
	N	360	0.8	–	Venturi/spray tower	3	15&18	60	10/75	1
	N	360	0.8	–	Venturi/spray tower	3	15 & 18	60	7/76	1
	R	125	0.5-1	Mechanical	Venturi/sieve tray tower	1	–	85	4/74	1
	R	125	0.5-1	Mechanical	Venturi/sieve tray tower	1	–	85	4/74	1
	N	125	0.5-1	Mechanical	Venturi/sieve tray tower	1	–	85	7/76	1
ineers--	R	120	2.5	ESP/ Venturi	Venturi rod absorber	3	–	90	9/75	1
rgas–Allied	R	115	3.2-3.5	ESP/ Venturi	Absorption tower	1	–	>90	8/77	5

ing in 1977